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Synthetic Fuels

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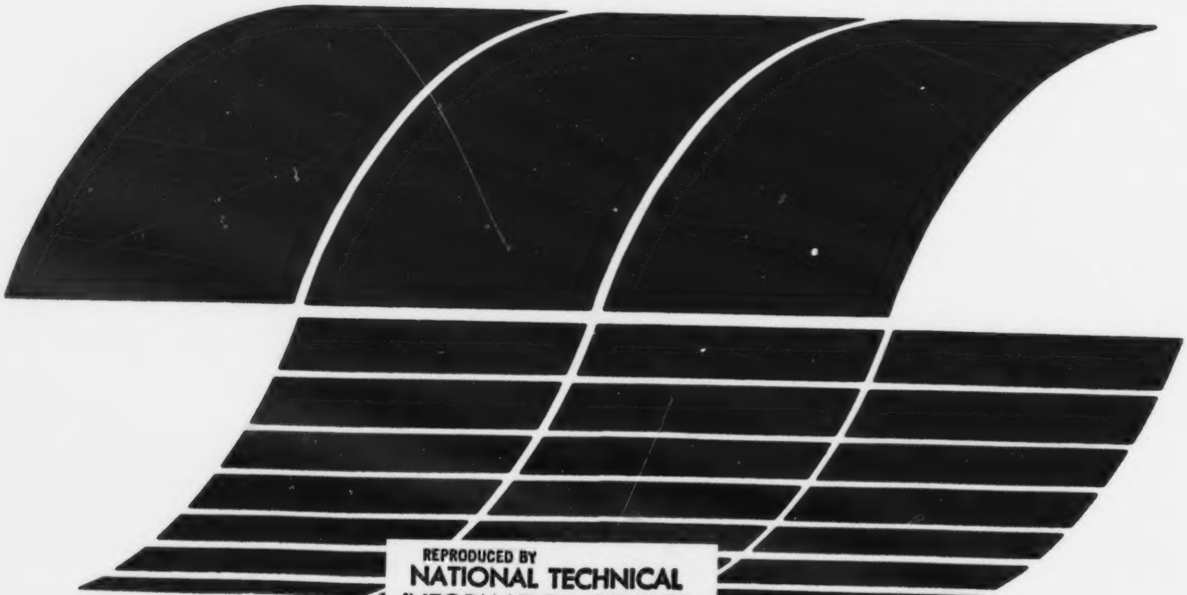

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16. ABSTRACT The report, part of a series to aid in determining the technological basis for New Source Performance Standards for Industrial Boilers, addresses the use of synthetic fuels produced from coal as a precombustion emission control for new industrial boilers. The synthetic fuels technologies considered include coal gasification and liquefaction. Examining the reduction of SOx, NOx, and particulate emissions in industrial boiler flue gases is emphasized. Two low-Btu gasification systems (the Wellman-Galusha gasifier with either the Stretford (W-G/S) or monoethanolamine (W-G/MEA) acid gas removal process) were selected for the detailed analyses, which involved two coal feedstocks (low-sulfur western and high-sulfur eastern) and five boiler capacities (8.8, 22, 44, 58.6, and 117 MWT heat input). The low-Btu gasification systems met the most stringent target NOx and particulate emissions control levels that were considered. For SO2 emissions, the W-G/S systems using low-sulfur coal achieved a stringent target emission control level of 43 ng SO2/J heat level. Using high-sulfur coal, the W-G/S system achieved a moderate target control level of 150 ng SO2/J heat input. For regulatory purposes, this assessment must be viewed as preliminary, pending results of the more extensive examination of impacts called for under Section 111 of the Clean Air Act.

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PREFACE

The 1977 Amendments to the Clean Air Act required that emission standards be developed for fossil-fuel-fired steam generators. Accordingly, the U.S. Environmental Protection Agency (EPA) recently promulgated revisions to the 1971 new source performance standard (NSPS) for electric utility steam generating units. Further, EPA has undertaken a study of industrial boilers with the intent of proposing a NSPS for this category of sources. The study is being directed by EPA's Office of Air Quality Planning and Standards, and technical support is being provided by EPA's Office of Research and Development. As part of this support, the Industrial Environmental Research Laboratory at Research Triangle Park, N.C., prepared a series of technology assessment reports to aid in determining the technological basis for the NSPS for industrial boilers. This report is part of that series. The complete report series is listed below:

Title	Report No.
The Population and Characteristics of Industrial/Commercial Boilers	EPA-600/7-79-178a
Technology Assessment Report for Industrial Boiler Applications: Oil Cleaning	EPA-600/7-79-178b
Technology Assessment Report for Industrial Boiler Applications: Coal Cleaning and Low Sulfur Coal	EPA-600/7-79-178c
Technology Assessment Report for Industrial Boiler Applications: Synthetic Fuels	EPA-600/7-79-178d
Technology Assessment Report for Industrial Boiler Applications: Fluidized-Bed Combustion	EPA-600/7-79-178e
Technology Assessment Report for Industrial Boiler Applications: NO_x Combustion Modification	EPA-600/7-79-178f
Technology Assessment Report for Industrial Boiler Applications: NO_x Flue Gas Treatment	EPA-600/7-79-178g
Technology Assessment Report for Industrial Boiler Applications: Particulate Collection	EPA-600/7-79-178h
Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization	EPA-600/7-79-178i

These reports will be integrated along with other information in the document, "Industrial Boilers - Background Information for Proposed Standards", which will be issued by the Office of Air Quality Planning and Standards.

ABSTRACT

This report is part of a series of technology assessment reports prepared for the EPA to aid in their determining the technological basis for new source performance standards for industrial boilers. The report addresses the use of synthetic fuels produced from coal as precombustion emission controls for new industrial boilers. The synthetic fuels technologies considered include coal gasification (low-, medium-, and high-Btu) and liquefaction. Major emphasis is placed on examining the reduction of SO_x, NO_x, and particulate emissions in the industrial boiler flue gases.

Based on detailed analyses of costs, energy, and environmental impacts, low-Btu coal gasification was selected as the "best synthetic fuels from coal" emission control technique for industrial boilers. Two low-Btu gasification systems, the Wellman-Galusha gasifier with either the Stretford (W-G/S) or monoethanolamine (W-G/MEA) acid gas removal process, were selected for the detailed analyses. Also, in performing the detailed analyses, two coal feedstocks (low-sulfur western and high-sulfur eastern) and five boiler capacities (8.8, 22, 44, 58.6, and 117 MW_T heat input) were considered.

All of the low-Btu gasification systems examined were capable of meeting the most stringent target NO_x and particulate emissions control levels considered (86 ng NO_x/J and 4 ng particulates/J heat input to the boiler). With respect to SO_2 emissions, the W-G/S systems using a low-sulfur coal feed could achieve a stringent target emission control level of 43 ng SO_2/J heat input. Using a high-sulfur coal feed, the W-G/S systems could achieve a moderate target control level of 150 ng SO_2/J heat input. The W-G/MEA systems were only considered for the high-sulfur coal cases and could be designed to achieve either the moderate or stringent target SO_2 control level.

The results of the costs analyses indicated that the annualized costs of the gasification/low-Btu gas-fired boiler systems were approximately 20 to 170 percent greater than the annualized costs of an equivalent capacity (on a boiler heat input basis) direct coal-fired boiler without pollution controls. The percentage increase in incremental costs varied indirectly with boiler capacity. For a given boiler capacity, the W-G/S system using low-sulfur coal had the lowest incremental costs, while the W-G/MEA system using high-sulfur coal had the highest incremental costs. All of the low-Btu gasification/boiler systems consume more energy (40-65%) than equivalent capacity uncontrolled coal-fired boilers. The low-Btu gasification systems are also sources of gaseous, liquid, and solid discharges. However, there do not appear to be any uncontrollable adverse environmental impacts. For regulatory purposes, this assessment must be viewed as preliminary, pending the results of the more extensive examination of impacts called for under Section III of the Clean Air Act.

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SECTION I

EXECUTIVE SUMMARY

1.1 INTRODUCTION

1.1.1 Background and Objective

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Specifically mentioned in the Act are fossil fuel-fired steam generators. Accordingly, EPA has undertaken a study of industrial boilers with intent to propose standards of performance based upon the results of this and other studies.

There are many ways to reduce emissions from fossil fuel-fired industrial boilers. These methods or techniques can be broadly classified as:

- Precombustion techniques
- Combustion techniques
- Post-combustion techniques.

This report presents the results of a study conducted to evaluate the use of synthetic fuels from coal technologies as precombustion emission controls for new industrial boilers. The synthetic fuels technologies examined included coal gasification (low-, medium-, and high-Btu) and liquefaction. Major emphasis was placed on the reduction of SO_x, NO_x, and particulate emissions in the industrial boiler flue gases. However, other gaseous pollutants in the flue gases, as well as gaseous emissions, liquid effluents, and solid wastes from the synthetic fuels systems, were also examined.

The results of this evaluation will be used by the EPA to assess the need for, and, if appropriate, to prepare a new source performance standard for small industrial boilers. Factors that were considered in evaluating the synthetic fuels technologies include: development status, capital and operating costs, energy impacts, environmental impacts, and performance data.

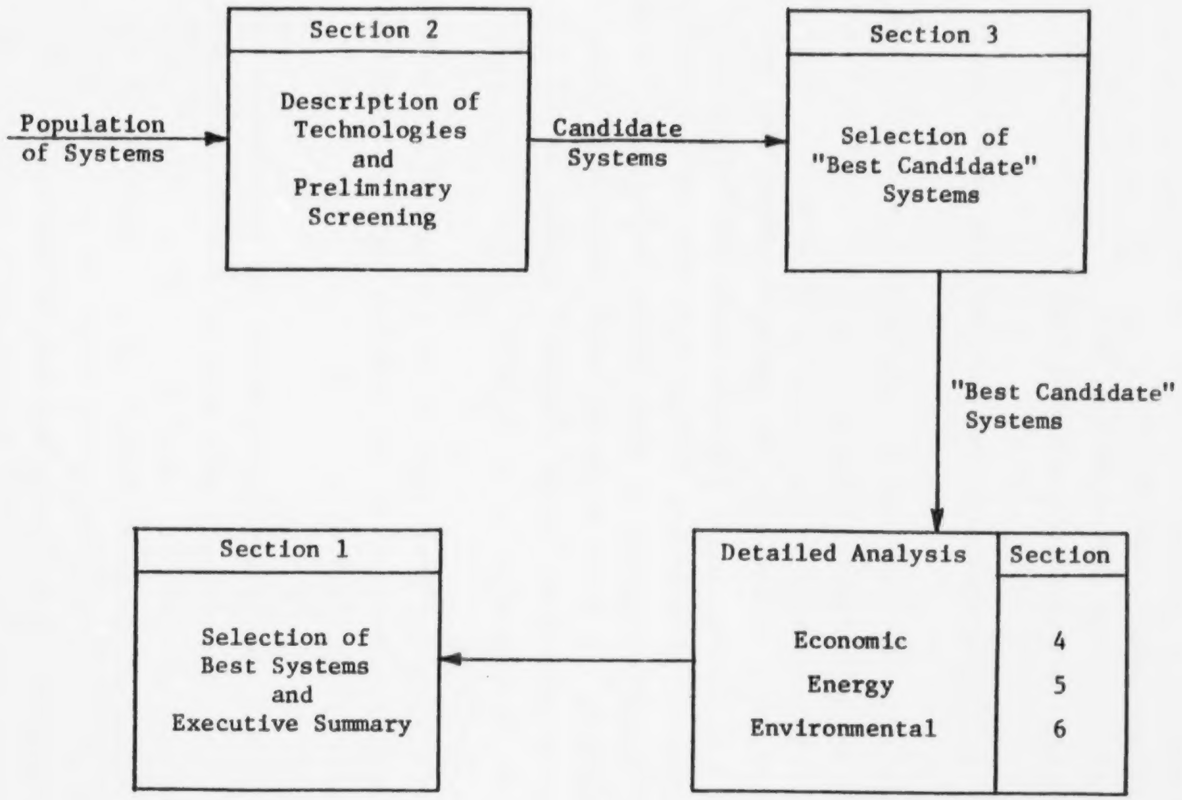
1.1.2 Approach and Report Organization

A multiphased approach was used to evaluate the synthetic fuels technologies as shown in Figure 1.1-1. First, the population of synfuels systems and their important components were identified. Because of the large number of systems that were considered, general type descriptions were prepared for each technology detailing items such as:

- Status of development,
- Applicability and limitations,
- Factors affecting performance, and
- System performance.

Using status of development as the major criterion, a preliminary screening of the synfuels population was performed in order to select several candidate systems for further study. As indicated in Figure 1.1-1, the systems description and preliminary screening step are addressed in Section 2 of this report.

The candidate synfuels systems identified in Section 2 were screened a second time in order to select 1 or 2 "best candidate" systems for detailed analysis. The criteria used were:



Section 2
Description of Technologies and Preliminary Screening
Population of Systems
Candidate Systems
Section 3
Selection of "Best Candidate" Systems
"Best Candidate" Systems
Detailed Analysis
Section
Economic
4
Energy
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Section 1
Selection of Best Systems and Executive Summary

Figure 1.1-1. Schematic of study approach.

- System performance,
- Status of development,
- Applicability to industrial boilers,
- Preliminary cost considerations,
- Preliminary energy considerations, and
- Preliminary environmental considerations.

The results of the second screening step are presented in Section 3.

The next step in the study approach was to perform detailed analyses of costs, energy, and environmental impacts for the "best candidate" systems. The results of those analyses are presented in Sections 4, 5, and 6 respectively. In performing the analyses, the synfuels systems were considered for application to several boiler sizes and fuels (see Table 1.1-1). The EPA had indicated at the beginning of this study that the coal-derived synfuels systems should be compared to other control technologies for coal-fired boilers. Therefore, the boiler sizes and fuels indicated in Table 1.1-1 correspond to the coal-fired "standard boilers" examined by the other EPA contractors for the Industrial Boiler Program.

The final step in the study approach was to identify the available source test data for combustion of the synthetic fuels in industrial boilers. That information is contained in Section 7.

In the ensuing discussion of emission control technologies, candidate technologies were compared using three emission control levels labelled "moderate, intermediate, and stringent" (see Table 1.1-2). These control levels were chosen only to encompass all candidate technologies and form bases for comparison of technologies for control of specific pollutants considering performance, costs, energy, and non-air environmental effects. From these comparisons, candidate "best" technologies (systems) for control of individual pollutants are recommended for consideration in subsequent industrial boiler studies. The systems selected and a discussion of their performance and impacts are presented in Section 1.2. These "best technology"

TABLE 1.1-1. STUDY BASIS BOILER SIZES AND COAL FEEDS

	Base Coal Analyses		
Boiler Heat Input, MW_T (10^6 Btu/hr)	Composition, Wt %	Eastern High Sulfur	Western Low Sulfur
8.8 (30)	Carbon	64.80	57.60
22 (75)	Hydrogen	4.43	3.20
44 (150)	Sulfur	3.54	0.60
58.6 (200)	Oxygen	6.56	11.20
117.2 (400)*	Nitrogen	1.30	1.20
	Moisture	8.79	20.80
	Ash	10.58	5.40
	Heating Value, MJ/kg (Btu/lb)	27.4 (11,800)	22.3 (9,600)

*This boiler size was examined only for a limited number of cases using low-Btu gasification.

TABLE 1.1-2. TARGET EMISSION CONTROL LEVELS FOR SYNTHETIC FUELS FROM COAL TECHNOLOGIES

	SO_2, ng/J (lb/10^6 Btu)	NO_x, ng (as NO_2)/J (lb/10^6 Btu)	Particulates, ng/J (lb/10^6 Btu)
Coal Gasification			
Moderate Control	150 (0.5)	86 (0.2)	13 (0.03)
Intermediate Control	86 (0.2)		
Stringent Control	43 (0.1)		
Coal Liquefaction			
Moderate Control	520 (1.2)	300 (0.7	13 (0.03)
Intermediate Control	260 (0.6)	220 (0.5)	
Stringent Control	86 (0.2)	86 (0.2)	

recommendations do not consider combinations of control technologies and have not undergone the detailed environmental, cost, and energy impact assessments necessary for regulatory action. Therefore, the levels of "moderate, intermediate, and stringent" and the recommendation of "best technology" are not to be construed as indicative of the regulations that will be developed for industrial boilers. EPA will perform rigorous examination of several comprehensive regulatory options before any decisions are made regarding the standards for emissions from industrial boilers.

1.2 SUMMARY OF BEST EMISSION REDUCTION SYSTEMS

In coal gasification systems the most important factor with respect to the system's performance as a precombustion emission control technique is the type of acid gas removal (AGR) process employed. Thus, for the gasification technologies it was necessary to examine both gasifiers and AGR processes. Approximately 70 low- and medium-Btu coal gasifiers and 70 AGR processes were considered in Section 2. From these populations, 14 gasifiers and 14 AGR processes were selected for additional consideration (see Tables 1.2-1 and 1.2-2). High-Btu coal gasification was eliminated from consideration as an emission control technique for industrial boilers. This was because 1) medium-Btu gas is just as applicable for use in industrial boilers as high-Btu gas, 2) high-Btu gasification is merely an extension of medium-Btu gasification (shift and methanation processing required in addition to medium-Btu gasification processing), and 3) high-Btu gasification has higher energy and cost impacts than medium-Btu gasification. Over twenty coal liquefaction processes were identified in Section 2. From this population four processes (see Table 1.2-3) were selected as candidates for additional consideration.

From the comparative evaluation of the systems listed in Tables 1.2-1 through 1.2-3, "best candidate" systems were selected for detailed analysis. These "best candidate" systems are listed in Table 1.2-4. A brief description of these systems is also contained in Table 1.2-4.

TABLE 1.2-1. COMPARISON OF CANDIDATE GASIFIERS

Gasifier	Type of Gas	Pressure	Development Status	Feedstock Restrictions	Environmental Impacts		Cold Gas Efficiency (%)[a]
					Direct	Indirect	
BGC Slagging Lurgi	Medium-Btu	High Pressure	Demonstration plant (started 1976)	Accepts all types; strongly caking requires agitation; 13-51 mm size coal; <20% moisture; crushing and sizing required; low ash coal may require fluxing agents.	Coal lock gas Slag quench vent gas Slag lock gas Slag quench blowdown Slag slurry	Waste streams from gas purification; process condensate quenching liquor, coal fines.	83
Bi-Gas	Low- or Medium-Btu	High Pressure	Not been demonstrated; pilot plant started in August 1976	Accepts all coal types; 70% less than 0.074 mm; crushing and pulverizing required; also slurry preparation or fluxing agent.	Slurry preparation vent gas Slag lock gas Slag slurry (solid & liquid)	Essentially no tars, oils, naphthas, or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	69
Chapman (Wilputte)	Low- or Medium-Btu	Atmospheric	Commercially available (since 1945) for Low-Btu gas; not commercially demonstrated for medium-Btu gas	Accepts all coal types; size <102 mm; crushing and sizing required.	Barrel valve vent gas Poke hole gases Ash pan gas Ash	Waste streams from gas purification; process condensate, quenching liquor; coal fines.	DNA
Coalex	Low-Btu	Atmospheric	Pilot plant since 1976	Accepts all coal types (lignite not been tested); <0.07 mm sized coal; crushing and pulverizing required; additives required.	Additive hopper vent gas Slag slurry (solid & liquid)	Essentially no tars or oils in raw product gas; gas purification may create additional unwanted streams.	DNA
Foster Wheeler/Stoic	Low-Btu	Atmospheric	Demonstration plant construction begun in 1977	Lignite, subbituminous, non-caking bituminous; 19-38 mm sized coal; crushing and sizing required; partial oxidation may be needed for strongly caking coals.	Coal hopper vent gas Ash pan gas Ash Poke hole gases	Waste streams from gas purification; tars/oils, coal dust.	77
GFERC Slagging	Medium-Btu	High Pressure	Pilot plant (1958-1965)	bituminous char, lignite char, lignite; 6.4-19 mm sized coal; <35% moisture; Crushing and sizing required.	Coal lock gas Slag quench vent gas Slag quench blowdown Slag slurry	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	85
Koppers-Totzek	Medium-Btu	Atmospheric	Commercially available since 1952	Accepts all coal types; 70% to 90% less than 0.074 mm; pulverizing required; ~1-8% moisture; possible addition of fluxing agents to lower ash fusion temp.	Coal bin N_2 vent Slag Slag slurry	Essentially no tars, oil, naphthas or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	75

(Continued)

TABLE 1.2-1. Continued

Gasifier	Type of Gas	Pressure	Development Status	Feedstock Restrictions	Environmental Impacts		Cold Gas Efficiency (%)[a]
					Direct	Indirect	
Lurgi	Low- or Medium-Btu	High Pressure	Commercially available (since 1941)	Accepts all coal; strongly caking coals may need partial oxidation or agitation; crushing and sizing required; <35% moisture; 3.2-38.1 mm sized coal.	Coal lock gas Ash lock gas Ash Ash quench water	Waste streams from gas purification; process condensate, quenching liquor, tars, oil, phenols, NH_3.	63-80
Pressurized Wellman-Galusha (MERC)	Low- or Medium-Btu	High Pressure	Pilot plant (since 1958); not commercially demonstrated for medium-Btu	Accepts all types; 50% <12.7 mm sized; no predrying; crushing and sizing required.	Coal lock gas Ash lock gas Ash	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	79
Riley-Morgan	Low- or Medium-Btu	Atmospheric	Commercially available for low-Btu gas (Pilot plant since 1975); not commercially demonstrated for medium-Btu gas	Accepts all coal types; 3.2-51 mm sized coal; crushing and sizing required.	Coal lock gas Ash pan gas Ash Poke hole gases	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	64-68
Texaco	Medium-Btu	High Pressure	Pilot plant	Accepts all coal types; 70% less than 0.074 mm; crushing, pulverizing, slurry preparation required.	Slurry preparation vent gas Slurry steam purge Preheater flue gases Slag slurry	Essentially no tars, oils, naphthas, or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	77
Wellman-Galusha	Low-Btu	Atmospheric	Commercially available (since 1941)	Can use anthracite, bituminous, charcoal, or coke; 7.9-14.3 mm for anthracite; 26-51 mm for bituminous; crushing and sizing required.	Coal bin gas Ash hopper gas Ash Poke hole gases	Waste streams from gas purification; coal fines, process condensate, quenching liquor.	75
Winkler	Low- or Medium-Btu	Atmospheric	Commercially available since 1926	Lignite, subbituminous, weakly caking bituminous; <9.53 mm sized coal; crushing required; <30% moisture for lignites; <18% moisture for higher rank coals; partial oxidation may be required.	Coal bin N_2 vent Dry ash bin N_2 vent Ash slurry settler vent Dry ash Ash slurry	Essentially no tars, oils, or naphthas in raw product gas; treatment produces process condensate and quenching liquor.	55-72
Woodall-Duckham/Gas Integrale	Low-Btu	Atmospheric	Commercially available (since 1940)	Lignite, bituminous; 6.4-38.1 mm sized coal; crushing and sizing required; drying not required; partial oxidation for strongly caking coals.	Coal hopper vent gas Ash lock gas Ash	Waste streams from gas purification; tar, process condensate, dust.	77

[a] $\text{Cold Gas Efficiency} = \dfrac{\text{Product Gas Energy Output}}{\text{Coal Energy Input}}$

DNA: Data not available

TABLE 1.2-2. CANDIDATE ACID GAS REMOVAL PROCESSES FOR GASIFICATION SYSTEMS

Acid Gas Removal Process	Description
Physical Solvent Processes	Physical solvent processes remove acid gases from the raw product gas by physical absorption in an organic solvent. They are most suited for high pressure applications because the solubilities of acid gases in the solvents are directly proportional to the acid gas partial pressure. Most of the solvents used in these processes have an appreciably higher affinity for H_2S than for CO_2, and can, therefore, be used in a manner that allows for selective removal of H_2S.
Estasolvan	
Fluor Solvent	
Purisol	
Rectisol	
Selexol	
Chemical Solvent Processes	Chemical solvent processes remove acid gases by adsorption followed by formation of a chemical complex between the acid gas and solvent. In most of these processes the solvent is regenerated by thermal decomposition of the chemical complex.
Benfield	
DEA	
DGA	
DIPA	
MDEA	
MEA	
Combination Physical/ Chemical Processes	Combination chemical/physical solvent processes use a physical solvent together with an alkanolamine chemical solvent additive. Both the chemical and the physical solvents remove CO_2, H_2S, and HCN; but the physical solvent is primarily responsible for the removal of acid gases such as CS_2, mercaptans, and COS.
Amisol	
Sulfinol	
Direct Conversion Process	The Stretford process is a liquid phase direct conversion process. H_2S is removed by adsorption in an alkaline solution, with subsequent oxidation of the sorbed H_2S to elemental sulfur. Organic sulfur compounds are not removed by the Stretford process.
Stretford	

TABLE 1.2-3. CANDIDATE COAL LIQUEFACTION SYSTEMS

Characteristics	SRC-I	SRC-II	H-Coal	Exxon Donor Solvent (EDS)
Status of Development	• 1700 kg/hr pilot plant operating, Ft. Lewis, WA • 210 kg/hr pilot plant, Wilsonville, AL	• 1700 kg/hr pilot plant operating, Ft. Lewis, WA • 210 metric ton/hr demonstration plant planned, West Virginia	• 100 kg/hr PDU operating, Trenton, NJ • 20 metric ton/hr pilot plant under construction, Cattlesburg, KY. Start-up in early 1979	• 0.9 and 2.4 kg/hr bench scale units and 34 kg/hr pilot plant operating, Baytown, TX • 8.6 metric ton/hr under construction, Baytown, TX Operation in 1980
Operational and Maintenance Requirements	• Solids accumulation control in the dissolver on western coals • Solids-liquid separation (filtration) equipment • Slurry pumps	• Solids accumulation control in the dissolver on western coals • Vacuum distillation column • Slurry pumps	• Ebullated bed reactor • Circulating slurry pumps	• Circulating slurry pumps • Flexicoking
Applicability	• Solid fuel produced can only be used in pulverized coal boiler	• Liquid fuel can be used in conventional fuel oil-fired boiler	• Liquid fuel can be used in conventional fuel oil-fired boiler	• Liquid fuel can be used conventional fuel oil-fired boiler
Environmental Impacts*	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes
Financial impact (capital cost of the liquefaction plant is prorated to the boiler size)	Capital cost may be as high as 50 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.
Energy Impact**	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%

*See Table 3.3-3, 3.3-4, and 3.3-5.

**Thermal Efficiency is defined as the energy content of all useful products expressed as a percentage of the energy content of the feed coal.

TABLE 1.2-4. "BEST CANDIDATE" SYNTHETIC FUELS SYSTEMS

"Best Candidate" Systems	Description
Low-Btu Coal Gasification	
Wellman-Galusha gasifier with Stretford AGR process	System operates at nominally atmosphere pressure; utilizes an alkaline solution to remove H_2S but does not remove COS; does not require a separate sulfur recovery process; capable of achieving stringent control levels for NO_x and particulates, stringent SO_2 level for low sulfur coal feed, and moderate SO_2 control level for high sulfur coal feed.
Wellman-Galusha gasifier with monoethanolamine (MEA) AGR process	System operates at nominally atmospheric pressure, although gas is moderately compressed ($\sim$0.3 MPa) for AGR process; requires separate sulfur recovery unit for AGR process off-gas (Claus/SCOT units); capable of achieving stringent control levels for NO_x, SO_x, and particulate emissions.
Medium-Btu Coal Gasification	
Lurgi gasifier with Stretford AGR process	High pressure operation ($\geq$2.4 MPa); utilizes alkaline solution to remove H_2S but does not remove COS; does not require separate sulfur recovery process; capable of achieving stringent control levels for NO_x and particulates, stringent SO_2 level for low sulfur coal feed, and intermediate SO_2 control level for high sulfur coal feed.
Lurgi gasifier with Rectisol AGR process	High pressure operation ($\geq$2.4 MPa); utilizes cold methanol for essentially complete removal of sulfur species; can meet stringent control levels for SO_x, NO_x, and particulate emissions.
Coal Liquefaction	
SRC-I	High pressure operation ($\sim$10 MPa); noncatalytic hydrogenation system; produces low melting point (155°C) solid fuel; capable of meeting stringent particulate control level with 80^+ percent efficiency particulate control device on flue gas; capable of meeting moderate SO_2 control level for high sulfur coal feed and intermediate SO_2 control level for low sulfur coal feed; NO_x emission reduction is questionable.
Exxon Donor Solvent (EDS)	High pressure operation ($\sim$14 MPa); donor solvent type hydrogenation system; produces distillate fuel oil type liquid product; includes liquids hydrotreating as part of process; capable of meeting stringent SO_2 (extra hydrotreating may be necessary for high sulfur coal feed) and particulate control levels; NO_x emission reduction is not well defined.

Based on the results of the detailed evaluations of the synfuels systems listed in Table 1.2-4, the two low-Btu gasification systems (Wellman-Galusha gasifier with either the Stretford or monoethanolamine AGR process) were selected as the best emission control systems for industrial boilers. Sections 1.2.1 through 1.2.7 summarize the results of the detailed analyses performed for those two systems. The following text briefly discusses why the "best candidate" medium-Btu gasification and liquefaction systems were not selected as the best systems.

There currently are no commercially operating medium-Btu gasification or liquefaction plants in the U.S. If and when built though, these plants are expected to be large capacity installations. Typical design capacities being considered are approximately 1760 MW_T (6000 x 10^6 Btu/hr) of medium-Btu gas output and approximately 3520 MW_T (12,000 x 10^6 Btu/hr or 50,000 bbl/SD) of coal liquids output. However, the industrial boilers examined in this study have relatively small fuel needs (up to 120 MW_T thermal heat input or 400 x 10^6 Btu/hr). Thus, an industrial boiler would require only a small percentage of a large synfuel plant's fuel production. Unless an unusual set of economic situations existed, a small industrial boiler operator probably would not (or could not), on his own, invest the capital necessary to build and operate a large medium-Btu coal gasification or liquefaction plant. The only way a small industrial boiler operator could utilize medium-Btu gas or coal liquids would be 1) to purchase a minor interest in a large plant being constructed by others, or 2) to purchase a portion of the fuel production from an existing plant. Both of those situations, however, require that someone already has built or wants to build the synfuels plant. Given these circumstances, medium-Btu coal gasification and liquefaction were not selected as the best emission control systems for industrial boilers.

1.2.1 Principles of Emissions Control

Converting coal into a "clean" low-Btu gas with subsequent combustion in a boiler, reduces SO_x, NO_x, and particulate emissions (versus direct coal combustion) by removing the pollutant's precursors. With respect to particulate emissions, the coal ash content is physically separated from the gas when the coal is gasified. Any entrained ash or coal particles are subsequently removed from the gas in the hot cyclones and gas quenching and cooling steps. SO_x emissions are reduced by removing sulfur species such as H_2S and COS from the low-Btu gas prior to combustion. Nitrogen oxide emissions are also reduced because low-Btu gas contains only small quantities of nitrogen compounds (NH_3 and cyanides) which can be oxidized to NO_x. Moreover, low-Btu gas burns with a low flame temperature which helps reduce the formation of NO_x by thermal fixation.

1.2.2 Applicability and Limitations to Boiler Size

The low-Btu gasification systems examined are applicable to any size industrial boiler. For the smallest industrial boilers examined (8.8 MW_T thermal input or 30 x 10^6 Btu/hr), one 3 m (10 ft) diameter Wellman-Galusha gasifier is required. For the larger boilers, multiple gasifiers are used (10 gasifiers are required for a boiler with a thermal input of 117.2 MW_T or 400 x 10^6 Btu/hr).

1.2.3 Emission Control Efficiency

There are no actual test data for the combustion of low-Btu gas in industrial boilers. However, SO_2 and particulate emission estimates based on engineering calculations should be good approximations for actual combustion results. NO_x emissions cannot be estimated quite as confidently. The mechanisms for NO_x formation are quite complex. Hence, an "a priori" estimate cannot be made.

As mentioned previously, the performance of synthetic fuels from coal systems as precombustion emission controls for industrial boilers was evaluated with respect to three target emission control levels (see Table 1.1-2). The target emission control levels attainable by the low-Btu gasification system examined are discussed in the following text.

For a low sulfur coal feed which produces relatively small amounts of organic sulfur compounds (∿50 ppmv), the Wellman-Galusha/Stretford (W-G/S) system can produce a fuel gas capable of meeting a stringent target SO_2 control level of 43 ng SO_2/J (0.1 lb/10^6 Btu). For a high sulfur coal feed, the COS concentration of the low-Btu gas is greater (estimated at ∿300 ppmv). In this situation, the W-G/S system can only achieve a moderate target SO_2 level of 150 ng/J (0.35 lb/10^6 Btu) since the Stretford process does not remove COS. However, the monoethanolamine (MEA) system can remove both COS and H_2S from the low-Btu gas. Thus, the Wellman-Galusha/MEA (W-G/MEA) system can produce a fuel gas from a high sulfur coal capable of meeting a stringent target SO_2 control level.

The MEA process can be designed to provide less complete removal of sulfur species if increased SO_2 emission levels are acceptable. Thus, the W-G/MEA system was examined for all three target SO_2 emission control levels. However, the H_2S removal reaction in the Stretford process is very fast, and residual levels are normally less than 100 ppmv. It is not possible to operate the process to achieve significantly less H_2S removal (although H_2S residuals around 100 ppmv may be possible with a less efficient type of contactor). For this reason, the W-G/S system was only examined for one target SO_2 emission control level (moderate control for high sulfur coal and stringent control for low sulfur coal).

For all of the cases examined, particulate emissions arising from combustion of the low-Btu gas are estimated to be similar to those for natural gas combustion--4 ng particulate/J (0.01 lb/10^6 Btu). This level is below the target particulate emission control level of 13 ng/J (0.03 lb/10^6 Btu).

NO_x emissions from combustion of any of the low-Btu gases examined are also estimated to be similar to those for natural gas combustion--50-100 ng NO_x/J (0.12-0.23 lb/10^6 Btu). This level is comparable to the target NO_x emission control level of 86 ng/J (0.2 lb/10^6 Btu). This estimated emission level is based on a limited number of tests conducted in small test furnaces.

In these tests, NO_x emissions were found to be similar to or less than those from natural gas combustion. The results were for combustion of low-Btu gases essentially free of NH_3 or cyanides. If present, those compounds can be oxidized to NO_x, increasing the overall quantity of NO_x formed.

1.2.4 Cost Impacts

The results of the economic evaluation of the low-Btu gasification/steam generation systems examined in this study are summarized in Table 1.2-5. Shown are the annualized costs of the low-Btu gas-fired boilers in dollars per year and dollars (per year) per unit of design heat input rate to the boiler. The difference (incremental costs) between these annualized costs and those for direct coal-fired boilers (without emission controls) are also listed in Table 1.2-5. The trend in the incremental annualized boiler costs is readily apparent when they are plotted as incremental costs per unit of heat input rate to the boiler versus heat input rate (see Figure 1.2-1). As shown in this figure, the low-Btu gasification/steam generation systems are more effective at higher heat input rates. This is due to realizing greater economies of scale in the low-Btu gasification/steam generation systems versus the direct coal-fired steam generation systems. Another result of the economic evaluation is that the annualized costs for the Wellman-Galusha/MEA systems do not vary discernibly with the level of SO_2 control.

At higher boiler heat input rates the gasification/steam generation systems are also more cost effective in controlling SO_2 emissions. This is illustrated in Figure 1.2-2 where the incremental annualized boiler costs per kg of SO_2 controlled (versus a direct coal-fired boiler) are plotted as a function of boiler heat input rate.

A complete cost component breakdown of the annualized costs for selected low-Btu gas-fired boilers is presented in Table 1.2-6. All of the synfuel cases examined are not included in this table since it is for illustrative purposes only. However, in all cases, the capital related charges and

TABLE 1.2-5. COST ANALYSIS OF LOW-BTU COAL GASIFICATION SYSTEMS FOR INDUSTRIAL BOILERS*

Boiler heat input, MW_T (10^6 Btu/hr)	Coal feed to gasifier	Sulfur control technique and level of control	SO_2 control efficiency, %	Annualized low-Btu gas-fired boiler costs: $\$10^3$/yr	Annualized low-Btu gas-fired boiler costs: per unit of heat input, $\$/kW_T$ ($\$/10^6$ Btu/hr)	Increase in costs over uncontrolled coal-fired boiler costs[b]: $\$10^3$/yr	Increase in costs over uncontrolled coal-fired boiler costs[b]: % of coal-fired
8.8 (30)	Low-sulfur	Stretford/Stringent	94.2[c]	2100	239 (70,000)	1120	115
8.8 (30)	High-sulfur	Stretford/Moderate	94.2[d]	2320	264 (77,200)	1370	144
8.8 (30)	High-sulfur	MEA/All Levels	94.2-98.2[e]	2540	289 (84,700)	1590	168
22 (75)	Low-sulfur	Stretford/Stringent	94.2[c]	2890	131 (38,500)	1020	54
22 (75)	High-sulfur	Stretford/Moderate	94.2[d]	3380	154 (45,100)	1530	83
22 (75)	High-sulfur	MEA/All Levels	94.2-98.2[e]	3870	176 (51,600)	2020	110
44 (150)	Low-sulfur	Stretford/Stringent	94.2[c]	4450	101 (29,700)	1330	42
44 (150)	High-sulfur	Stretford/Moderate	94.2[d]	5340	121 (35,600)	2260	73
44 (150)	High-sulfur	MEA/All Levels	94.2-98.2[e]	6050	138 (40,300)	2970	97
58.6 (200)	Low-sulfur	Stretford/Stringent	94.2[c]	5320	91 (26,600)	950	21
58.6 (200)	High-sulfur	Stretford/Moderate	94.2[d]	6530	111 (32,700)	2280	54
58.6 (200)	High-sulfur	MEA/All Levels	94.2-98.2[e]	7350	125 (36,800)	3100	74
117.2 (400)	Low-sulfur	Stretford/Stringent	94.2[c]	9340	80 (23,400)	1410	18
117.2 (400)	High-sulfur	Stretford/Moderate	94.2[d]	11,810	101 (29,700)	4030	53

[a] Wellman-Galusha gasifier; mid-1978 dollars; 60% annual operating factor.

[b] Low-Btu gas-fired costs were compared to uncontrolled coal-fired boiler costs for systems using the same coal feedstock and having identical boiler heat inputs.

[c] SO_2 emission level is 30 ng/J (0.07 lb/10^6 Btu).

[d] SO_2 emission level is 140 ng/J (0.32 lb/10^6 Btu).

[e] SO_2 emission levels are 140-43 ng/J (0.32-0.10 lb/10^6 Btu).

All systems have estimated NO_x and particulate emissions as follows:
NO_x--50-100 ng/J (0.12-0.23 lb/10^6 Btu)
particulates--<4 ng/J (<0.01 lb/10^6 Btu).

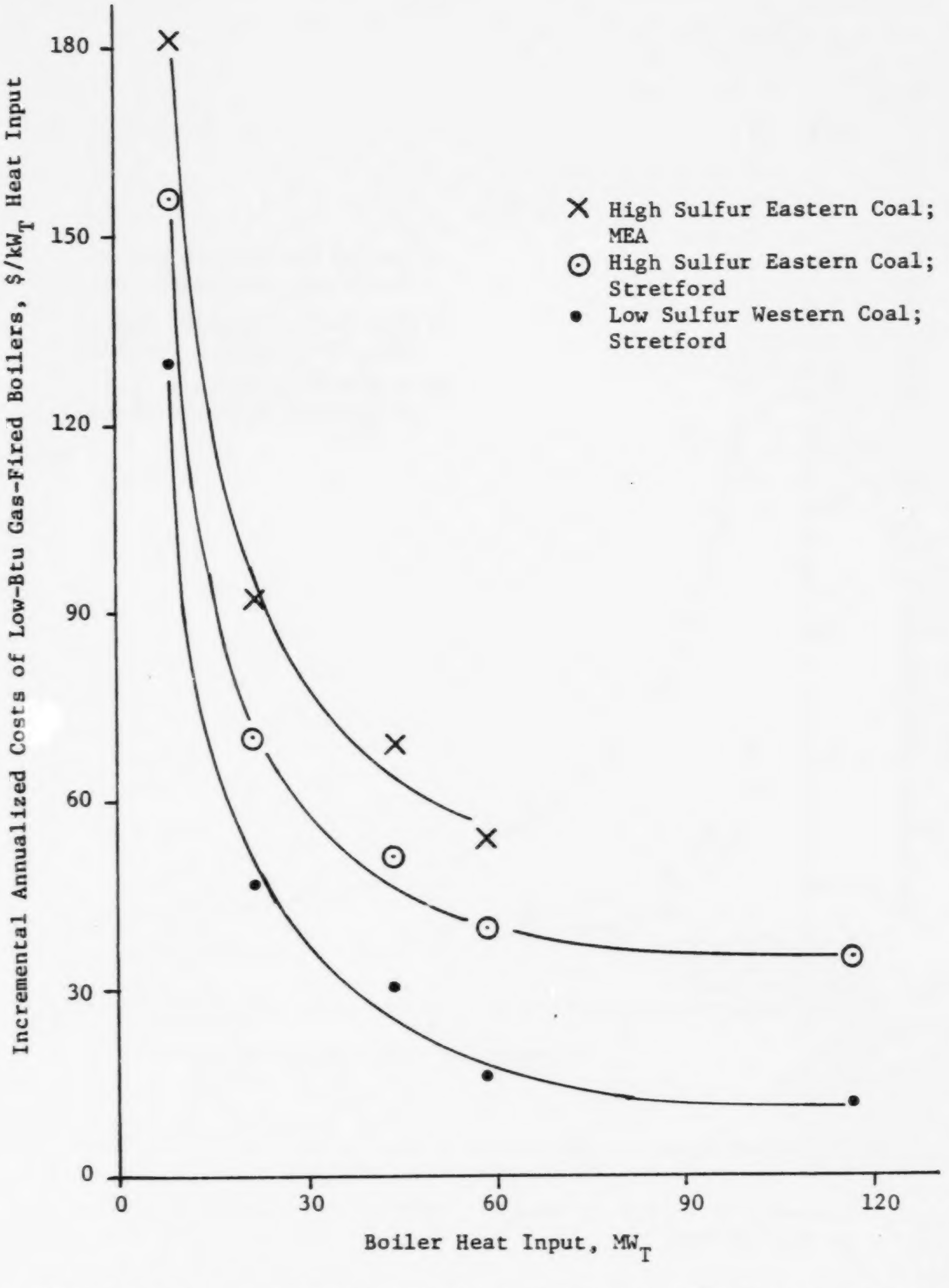
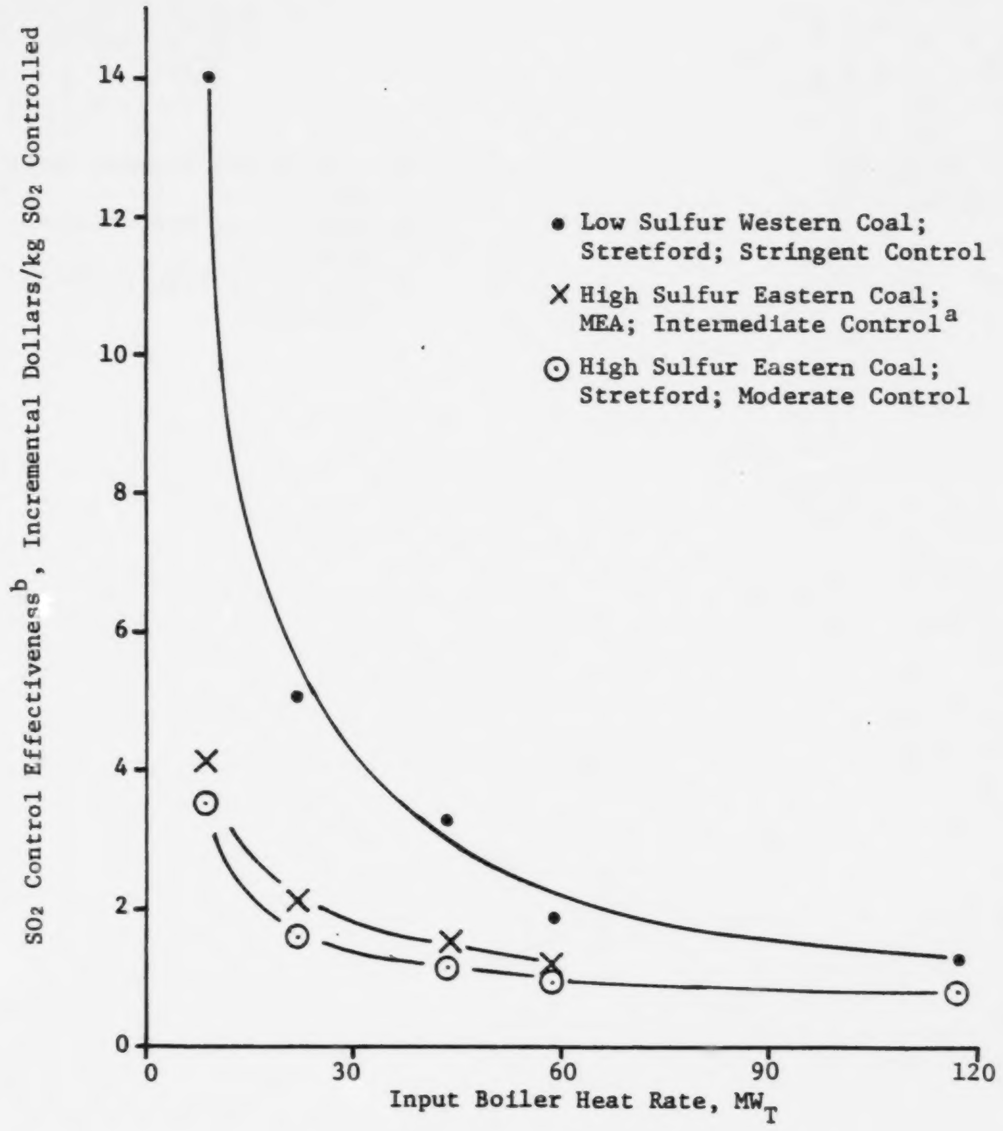


Figure 1.2-1. Incremental annualized costs of low-Btu gas-fired boilers versus boiler input heat rate.



[a] Values for Moderate and Stringent SO_2 Control have less than a 5% variation from the intermediate control value.

[b] Assumes that total incremental control costs are allocated to SO_2 removal.

Figure 1.2-2. Cost effectiveness of low-Btu gasification/steam generation systems in controlling SO_2 emissions.

TABLE 1.2-6. SENSITIVITY ANALYSIS--LOW-BTU GAS-FIRED BOILER ANNUALIZED COSTS

Coal	Low Sulfur Western		High Sulfur Eastern	
Boiler Input Heat Rate, MW_T	8.8	58.6	8.8	58.6
	Percent of Annualized Costs			
Labor and Maintenance	11.4	7.7	10.3	6.3
Utilities	1.4	1.1	1.3	0.9
Overhead	4.3	3.2	3.9	2.6
Capital Charges	5.2	5.8	4.7	4.7
Fuel (Low-Btu gas)				
Labor & Maintenance	23.8	18.2	22.4	15.6
Coal	4.3	11.3	6.9	16.5
Other Operating Costs	4.3	12.2	6.9	16.8
Overhead	9.0	6.8	8.6	5.8
Capital Charges	36.2	33.6	34.9	30.6
Total Fuel	77.6	82.1	79.7	85.3
Annualized Costs, 10^3\$/year	2100	5320	2320	6530

Basis: 1) A Stretford is the acid gas removal system for all cases shown.

2) The first four items in the table above - labor and maintenance, utilities, overhead, capital charges - refer to the gas-fired boiler system. The fifth item - fuel - is low Btu gas from the Wellman-Galusha System. Percentages under fuel are the various cost components of the gasification system.

operating and maintenance costs are the largest contributors to annualized costs. The coal cost is the next largest contributor.

The coal costs used in this study are typical of minemouth costs. Since some degree of coal transportation and handling is likely for most industrial applications, the effect of coal costs on annualized costs was evaluated. Moreover, since the coal-to-gas energy conversion in the gasification system is less than unity, an increase in coal costs will affect the annualized costs of a low-Btu gas-fired boiler to a greater extent than the annualized direct coal-fired boiler costs. Examples of the extent of this impact are presented in Table 1.2-7. For illustrative purposes, coal transportation costs were taken as \$1.00/GJ and \$0.50/GJ for the low sulfur western and high sulfur eastern coals, respectively.

1.2.5 Energy Impacts

The gasification of coal to produce a low-Btu gas incurs a significant energy penalty. For the Wellman-Galusha/Stretford (W-G/S) system gasifying low sulfur western coal, about 39 percent more energy is input to the system in the form of coal, steam, and electricity than is recovered in the product low-Btu gas. Thus, for direct coal-fired and low-Btu gas-fired boilers with the same heat input rate, a low-Btu gasification/boiler system will consume 39 percent more energy than the direct coal-fired boiler (excluding auxiliary energy requirements of the boilers). To gasify high sulfur eastern coal, the W-G/S system energy penalty is about 48 percent. The Wellman-Galusha/MEA (W-G/MEA) system using high sulfur eastern coal has an even higher energy penalty of about 64 percent. However, as was true for the economic impacts, the energy penalty of W-G/MEA is essentially unaffected by the SO_2 control level achieved.

The major contributor to the energy consumed by the low-Btu gasification systems is the gasification inefficiency. This includes both conversion losses and the energy content of the by-product tars and oils. Use of the

TABLE 1.2-7. EFFECT OF COAL TRANSPORTATION COSTS ON ANNUALIZED LOW-BTU GAS-FIRED BOILER COSTS[a]

	Coal Cost = Base Case Values[b]			Coal Transportation Costs Added[c]		
Input Heat Rate to Boiler, MW_T (10^6 Btu/hr)	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Low-Btu Gas-fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Low-Btu Gas-Fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr
Low Sulfur Western Coal						
8.8 (30)	980	2100	1120	1140	2330	1190
58.6 (200)	4370	5320	950	5480	6830	1350
High Sulfur Eastern Coal						
8.8 (30)	950	2320	1370	1040	2430	1390
58.6 (200)	4250	6530	2280	4800	7260	2460

[a] All gasification facilities use the Stretford process

[b] Base Case Values: Low Sulfur Western (LSW) - \$0.40/GJ; High Sulfur Eastern (HSE) - \$0.68/GJ

[c] Transportations Costs: LSW - \$1.00/GJ; HSE - \$0.50/GJ

by-products' energy would lower the energy penalties just presented by about 20 percentage points (e.g., from 64 to 44 percent for a W-G/MEA system gasifying high sulfur eastern coal).

1.2.6 Environmental Impacts

The low-Btu gasification systems examined are sources of gaseous, liquid, and solid discharges. However, there do not appear to be any uncontrollable adverse environmental impacts associated with the production and use of low-Btu gas. Tables 1.2-8, 1.2-9, and 1.2-10 summarize the air emissions, water effluents, and solid wastes from the various low-Btu gasification systems examined.

1.2.7 Commercial Availability

In the U.S. at the present time, there are no commercially operating low-Btu coal gasification systems which incorporate an acid gas removal (AGR) process for gas cleanup. However, the AGR processes considered in this study are commercially available from a number of process licensors and vendors. The ability of those vendors, as well as gasifier vendors, to meet an increased demand for their systems cannot be easily estimated. Most likely though, a rapid rise in demand would strain supplies, at least until additional fabrication capacity could be constructed. A more gradual rise in demand for low-Btu gasification facilities would probably have little impact on supplies since industry would have time to respond to the new market conditions.

TABLE 1.2-8. AIR POLLUTION IMPACTS FROM LOW-BTU GASIFICATION SYSTEMS[a]

Systems				Emissions						Other Pollutants	
				SO_2		NO_x		Particulates			
Coal Feed	SO_2 Control Level (% Reduction)	Emission Source	Sulfur Species Removal Process	ng/J	(lb/10^6Btu)	ng/J	(lb/10^6Btu)	ng/J	(lb/10^6Btu)	Pollutant	Change from Uncontrolled[b]
Low sulfur	Direct coal combustion (uncontrolled)	Combustion gas	-	520	(1.2)	340-400	(0.78-0.94)	600-1900	(1.4-4.5)		
Low sulfur	Stringent(94.2)	Combustion gas	Stretford	30	(0.07)	50-100	(0.12-0.23)	<4	(<0.01)	Organics CO Trace elements	-- - -
		Gasification system[c]		1.9	(0.005)	<0.4	(<0.001)	Small		Organics CO NH_3 HCN H_2S COS	+ + + + + +
High sulfur	Direct coal combustion (uncontrolled)	Combustion gas	-	2400	(5.7)	280-330	(0.64-0.76)	950-3100	(2.2-7.2)		
High sulfur	Moderate(94.2)	Combustion gas	Stretford	140	(0.32)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[c]		9.0	(0.02)	<0.4	(<0.001)	Small		Same as low sulfur	
	Moderate(94.2)	Combustion gas	MEA[e]	140	(0.32)	50-100)	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	
	Intermediate (96.5)	Combustion gas	MEA[e]	86	(0.2)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	
	Stringent(98.2)	Combustion gas	MEA[e]	43	(0.1)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	

[a] Wellman-Galusha gasifier with the indicated sulfur control processes.

[b] ++ increase; + slight increase; -- decrease; - slight decrease.

[c] Major emissions of SO_2, NO_x came from incinerated coal-feeder vent. Other emission sources are incinerated start-up gases, fugitives, and Stretford oxidizer vent.

[d] Major emissions of SO_2, NO_x came from incinerated coal-feeder vent, with smaller amounts from SCOT tail gas. Other emission sources are incinerated start-up gases and fugitives.

[e] Including Claus and SCOT for acid gas treatment.

TABLE 1.2-9. WATER POLLUTION IMPACTS FROM LOW-BTU GASIFICATION SYSTEMS[a]

Coal Feed	SO_2 Removal Level[b]	Sulfur Species Removal Process	Effluent Generated	
			Type	Quantity, ng/J ($lb/10^6$ Btu)
Low Sulfur	94.2%	Stretford	• coal pile runoff	• not quantified[c]
			• ash sluicing water	• not quantified[d]
			• process condensate	5600 (13)[e,f]
			• Stretford blowdown	300 (0.7)[f,g]
High Sulfur	94.2-98.2%	Stretford or MEA	• coal pile runoff	• not quantified[c]
			• ash sluicing water	• not quantified[d]
	94.2%	Stretford	• process condensate	1400 (3.4)[e,f]
			• Stretford blowdown	610 (1.4)[f,g]
	94.2-98.2%	MEA	• process condensate	9100 (21)[e,f]

---

[a] Wellman-Galusha gasification system.

[b] Effluents are the same for moderate, intermediate and stringent control levels.

[c] The quantity of water runoff from coal storage piles is largely site dependent; where coal is stored in small quantities, covered bins are usually employed, resulting in no water runoff from coal storage.

[d] The quantity of water used for ash sluicing varies widely, depending on the practices of the individual gasifier operators. The ash may be removed essentially dry, with the addition of little water. As a worst case, 1 kg of water may be used per kg of ash removed. This results in ash sluicing water of 3400-5600 ng/J (8-13 $lb/10^6$ Btu).

[e] The quantities of process condensate depends on the water content of the coal feed, the desired temperature of cooled product gas, and the pressure of the product gas. Systems featuring the MEA process produce a gas at 280 kPa (40 psia). Values shown are the quantities of condensate sent to on-site multiple effect evaporators. Residual wastes from the evaporators may be as little as 5% of the values shown.

[f] The estimated compositions of process condensate and Stretford blowdown are shown in Tables 6.1-3, 6.1-4, and 6.1-5.

[g] Based on: complete conversion of HCN in feed gas to NaCNS; 2% conversion of H_2S to $Na_2S_2O_2$; purge when salts concentration reaches 25% (Ref. 6-5, 6-8, 6-9).

TABLE 1.2-10. SOLID WASTE IMPACTS FROM LOW-BTU GASIFICATION SYSTEMS[a]

Coal Feed	SO_2 Removal Level (Control Efficiency, %)	Acid Gas Removal Process	Solid Waste Generated		
			Type	ng/J	(lb/10^6 Btu)
Low sulfur	Direct Coal Combustion (Uncontrolled)	-	Bottom ash	473-1810	(1.1-4.2)
Low sulfur	1[b] (94.2)	Stretford	Gasifier ash	3760	(8.73)
			Cyclone dust	420	(.977)
			Sulfur cake	508	(1.18)
High sulfur	Direct Coal Combustion (Uncontrolled)	-	Bottom ash	775-2880	(1.8-6.7)
High sulfur	Moderate (94.2)	MEA[c]	Gasifier ash	6540	(15.2)
			Cyclone dust	367	(.853)
			Sulfur	1650	(3.83)
			MEA blowdown	116	(0.27)
	Intermediate (96.5)	MEA[c]	Sulfur[d]	1680	(3.90)
			MEA blowdown	116	(0.27)
	Stringent (98.2)	MEA[c]	Sulfur[d]	1690	(3.93)
			MEA blowdown	116	(0.27)
	Moderate (94.2)	Stretford	Sulfur cake[d]	3300	(7.67)

[a] Wellman-Galusha gasification system.
[b] The same system is used for moderate, intermediate, or stringent.
[c] Including Claus to treat acid gas; SCOT to treat tail gas.
[d] Ash and dust are the same for all 3 control levels.

SECTION II.

DESCRIPTION OF SYNTHETIC FUELS FROM COAL SYSTEMS

The combustion of coal in an industrial boiler produces flue gases containing sulfur and nitrogen oxides and entrained particulates. The quantities of those pollutants depend on several factors including:

- the sulfur and ash content of the coal feed,
- the type of boiler used, and
- the heat input rate to the boiler.

The levels of uncontrolled SO_2, NO_x, and particulate emissions from coal-fired industrial boilers can be estimated using EPA emission factors (Ref. 2-1). Those estimated uncontrolled emissions (expressed as ng pollutant per Joule of heat input to the boiler or lb per 10^6 Btu input) are shown below for combustion of the high sulfur eastern and low sulfur western coals being examined in this study.

	High Sulfur Eastern Coal	Low Sulfur Western Coal
Coal Feed		
Wt. % Sulfur	3.54	0.6
Wt. % Ash	10.6	5.4
Higher Heating Value, MJ/kg (Btu/lb)	27.4 (11,800)	22.3 (9,600)
Flue Gas Components, ng/J (lb/10^6 Btu)		
Particulates		
Pulverized Coal	3080 (7.16)	1930 (4.48)
Spreader Stoker	2510 (5.83)	1570 (3.65)
All Other Stokers	960 (2.23)	600 (1.39)

	High Sulfur Eastern Coal	Low Sulfur Western Coal
Flue Gas Components, ng/J (lb/10^6 Btu)		
SO_2; all types	2420 (5.62)	510 (1.18)
NO_x; as NO_2		
Pulverized Coal	330 (0.767)	400 (0.930)
Stokers	270 (0.627)	340 (0.790)

Emission control techniques for industrial boilers using coal as the starting fuel can be divided into three categories:

- Pre-combustion controls
  - Synthetic fuels from coal
  - Physical or chemical coal cleaning

- Combustion controls
  - Combustion modifications

- Post-combustion controls
  - Particulate removal
  - Flue gas desulfurization
  - Flue gas treating for NO_x

This study addresses the use of synthetic fuels from coal as a precombustion emission control. The synfuels techniques examined are coal liquefaction and low-, medium-, and high-Btu coal gasification.

These coal conversion techniques reduce the formation of SO_2, NO_x and particulates by converting coal into "clean" fuels. During the conversion process, ash is removed and portions of the nitrogen and sulfur content of the coal react to form NH_3, HCN, H_2S, COS, etc. The nitrogen and sulfur compounds are then removed by commercially available processes. The quality of the final product from a gasification or liquefaction process depends on the purification and upgrading processes employed. It is possible to produce a coal-derived gaseous fuel that is essentially identical to natural gas (the least polluting fossil fuel currently available), or a coal-derived liquid fuel that is comparable to high-grade distillate fuel oil.

The approach used in assessing synthetic fuels from coal technologies as emission controls for industrial boilers was presented in Section I and shown schematically in Figure 1.1-1. The first step in the approach is to describe the technologies and select a set of candidate systems for further analysis (See Figure 2.1-1). The results of this first step are presented in this section. Section 2.1 deals with low- and medium-Btu coal gasification, Section 2.2 with high-Btu gasification, and Section 2.3 with coal liquefaction.

2.1 LOW- AND MEDIUM-BTU COAL GASIFICATION

Coal can be gasified with either air or pure (98+ percent) oxygen to produce synthetic gas. If air is the reactant, a low-Btu gas results; if pure oxygen is the reactant, a medium-Btu gas is produced. Since the only difference in a low- or medium-Btu system is the presence or absence of an oxygen production unit, these two types of gasification systems are described together in the following sections.

2.1.1 System Description

Coal gasification systems can be considered to consist of three basic process operations: coal pretreatment, coal gasification, and gas purification. Each of these operations consist of process modules which are employed to satisfy the functions of the operations. In addition, pollution control operations are required to handle the multimedia (air, water, and solid) effluents from the three process operations. The pollution control operations can be conveniently grouped into air pollution control, water pollution control, and solid waste management modules. The selection of specific

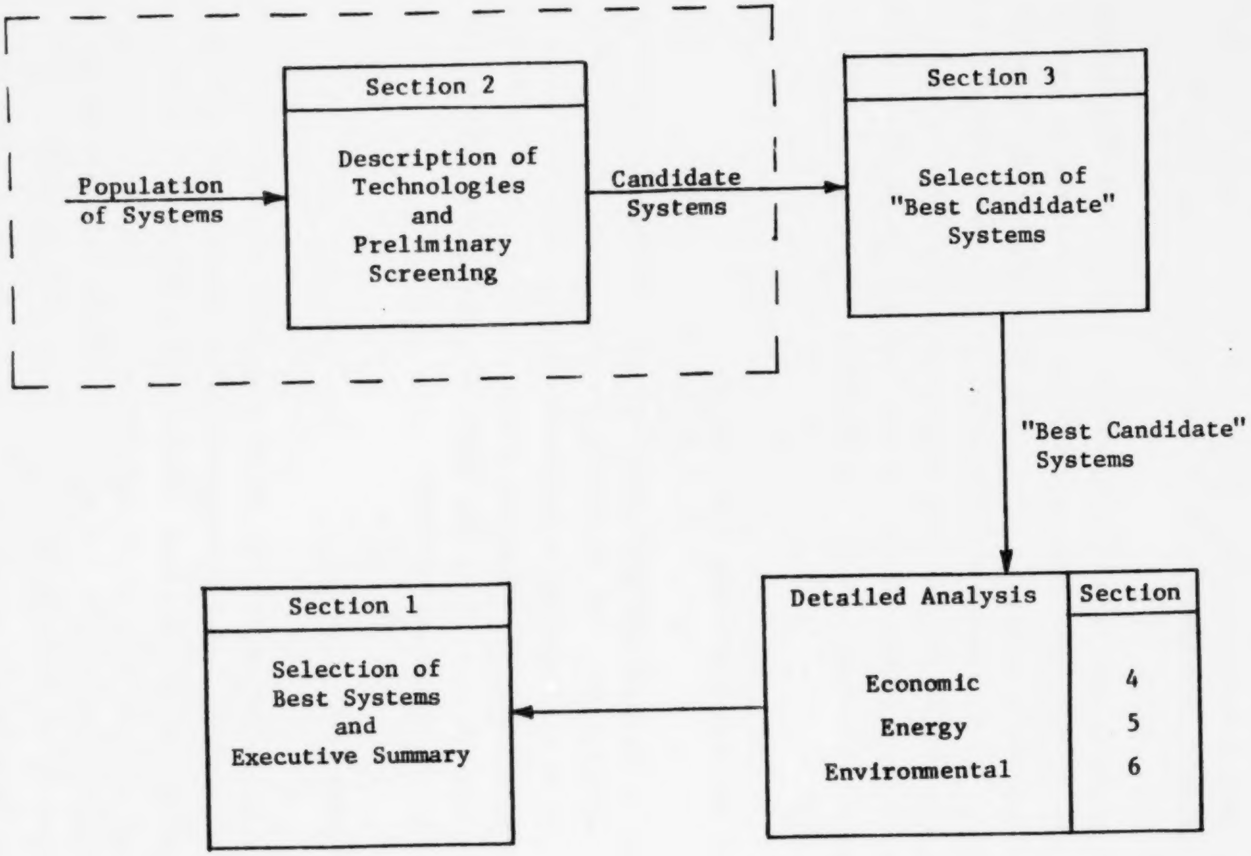

Section 2
Description of Technologies and Preliminary Screening
Population of Systems
Candidate Systems
Section 3
Selection of "Best Candidate" Systems
"Best Candidate" Systems
Detailed Analysis
Section
Economic
Energy
Environmental
4
5
6
Section 1
Selection of Best Systems and Executive Summary

Figure 2.1-1. Schematic of study approach.

processes for a process or pollution control module depends on the intended use of the low- or medium-Btu gas, the product specifications associated with that end use, and the requirements of the other processes within an integrated gasification system. The process modules and their general relationship within a system are depicted in Figure 2.1-2. The various pollution control modules which may be required in a gasification system are also listed in Figure 2.1-2.

In the following text, the purpose of the three basic operations and pollution control modules is discussed and a list of available processes is presented. Again, it is not possible to identify a process or set of processes to be used in a specific facility. Those are decisions that must be made when the facility is being designed, and will depend on plant location, the available coal feed, the desired product gas specifications, and economic and environmental considerations.

2.1.1.1 Coal Pretreatment--

The primary function of the coal pretreatment operation is to supply a coal feed which satisfies the physical specifications of the gasification operation. Coal handling and storage modules are generally required in this operation. The need for other processing modules such as grinding and screening depends on the nature of the coal feedstock and the requirements of the other coal pretreatment modules and the gasification operation. Coals having excess moisture may need to be dried while caking bituminous coals may need to be partially oxidized to reduce their caking tendencies. The feed coal must be crushed and sized for fixed-bed gasifiers or pulverized for fluid-bed or entrained-flow gasifiers. Fines from crushing processes which cannot be fed directly to a fixed-bed gasifier can either be sold as a by-product, consumed to supply on-site fuel needs, or briquetted and fed to the gasifier. Table 2.1-1 lists the functions and equipment used in each of the seven coal pretreatment process modules indicated in Figure 2.1-2.

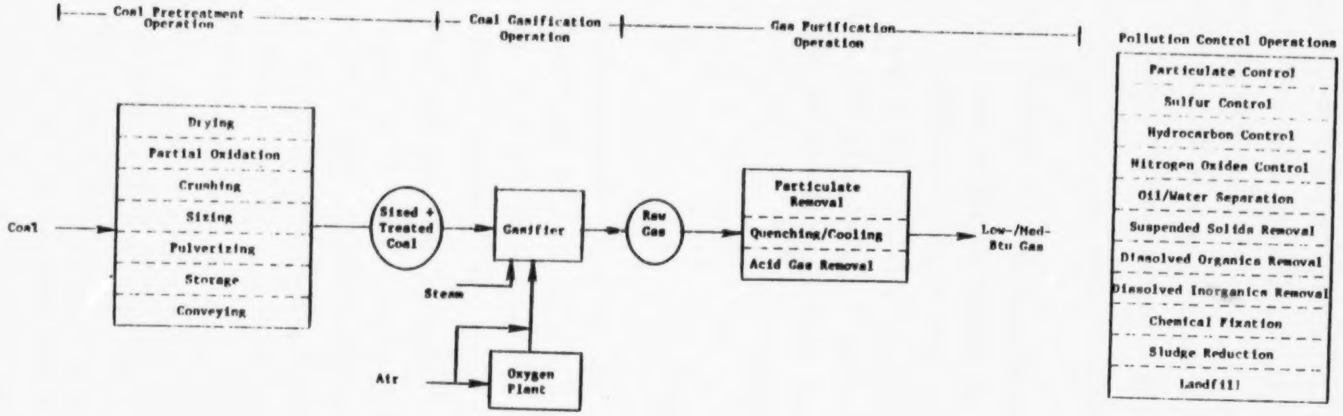

Coal Pretreatment Operation
Coal Gasification Operation
Gas Purification Operation
Pollution Control Operations
Coal
Drying
Partial Oxidation
Crushing
Sizing
Pulverizing
Storage
Conveying
Sized + Treated Coal
Gasifier
Steam
Air
Oxygen Plant
Raw Gas
Particulate Removal
Quenching/Cooling
Acid Gas Removal
Low-/Med-Btu Gas
Particulate Control
Sulfur Control
Hydrocarbon Control
Nitrogen Oxides Control
Oil/Water Separation
Suspended Solids Removal
Dissolved Organics Removal
Dissolved Inorganics Removal
Chemical Fixation
Sludge Reduction
Landfill

Figure 2.1-2. Low- or medium-Btu coal gasification system process and pollution control modules.

TABLE 2.1-1. FUNCTIONS OF MODULES IN COAL PRETREATMENT OPERATION

Module	Function	Equipment Used
Handling/ Transportation	Movement of coal to and from other pretreatment modules and to the gasification operation.	Belt conveyors Bucket elevators
Storage	Provides adequate reserves to allow for supply/demand surges (mine or gasification plant downtime) and possibly blending capability to provide a uniform feed to the gasification operation.	Covered/uncovered bins - up to about 1.8 x 10^6 kg (2000 short tons) per bin; uncovered piles on ground for greater than 2.3 x 10^8 kg (250,000 short tons)
Crushing/ Sizing	Size reduction and elimination of over- and undersize coal particles from a fixed-bed gasifier feed stream. Size specifications dictated by mechanical characteristics of gasifier.	Crushing - double and single roll crushers, rotary breakers, impactors, cage mills Sizing - Coarse (>50 mm [2 in.] particles) - grizzly screens Medium (>13 mm [½ in.] particles) - revolving, shaking or vibrating screens Fine (>2 mm [0.08 in.] particles) - oscillating screens
Briquetting	Compaction of coal fines to produce a briquette of a size suitable for feed to a fixed-bed gasifier. Certain binders, such as asphalt or tar, may be required, along with a baking or curing step, to give the briquette the required structural strength.	Coal fines hopper, feeder and either a rotating or a plate-type press along with provisions for adding a binder. A baking oven may also be required.
Pulverizing	Size reduction to provide a feedstock for a fluid- or entrained-bed gasifier.	Hammer mills Cage mills Impactors Ball mills
Drying	Mechanical dewatering or heat treatment to remove excess moisture from coal feed.	Mechanical: centrifugal; filtration Thermal: fixed- or fluid-bed driers.
Partial Oxidation	Method of achieving a reduction in the caking tendencies of a feed coal by contacting the coal with hot air or combustion gases under controlled conditions (temperature; time) in a suitable reactor.	Same as thermal driers.

Source: Ref. 2-2

2.1.1.2 Coal Gasification--

The function of the coal gasification operation is to produce a raw low- or medium-Btu gas by reacting coal with a steam/air or steam/oxygen mixture. If air is the source of oxygen, a low-Btu gas with a heating value of around 5.6 MJ/m^3 (150 Btu/scf) is produced. When essentially pure oxygen is used, a medium-Btu gas is produced with a heating value around 13.0 MJ/m^3 (350 Btu/scf). Table 2.1-2 shows ranges in raw product gas composition that are typical for coal-derived low- and medium-Btu gases.

TABLE 2.1-2. TYPICAL RAW PRODUCT GAS COMPOSITIONS FOR LOW- AND MEDIUM-BTU GASES

	Low-Btu Gas	Medium-Btu Gas
Composition, vol. % (dry)		
CO_2	3-14	6-31
CO	16-29	17-61
CH_4	1-5	1-14
H_2	11-23	23-39
N_2	40-63	1-3
H_2S, ppmv	400-6,900[a]	3,700-15,000[a]
NH_3, ppmv	b	b
HHV, MJ/m^3 (Btu/scf)	4.8-7.3 (130-200)	9.4-13.9 (250-370)

[a] H_2S concentration is proportional to the coal sulfur content. Range shown for medium-Btu gas and lower end of range for low-Btu gas are based on available data. Upper end of range for low-Btu gas is what might be expected from gasifying a 3.5 percent sulfur coal (based on engineering calculations).

[b] Amount dependent on coal nitrogen content and gasifier operating conditions. Data are not available to give a representative range.

Source: References 2-2 through 2-18

Numerous reactions occur when coal is gasified. Among the more important ones with respect to the major raw gas constituents are:

1) $C + \frac{1}{2}O_2 \rightarrow CO$

2) $H_2O + C \rightarrow CO + H_2$

3) $CO + H_2O \leftrightarrows CO_2 + H_2$

4) $2H_2 + C \rightarrow CH_4$

5) $CO + 3H_2 \leftrightarrows CH_4 + H_2O$

From the viewpoint of producing a clean fuel for use in an industrial boiler, reactions involving the sulfur and nitrogen present in coal are of greater interest than the reactions listed above. This is because these reactions result in the formation of a variety of compounds which, if not removed in downstream processing steps, can be oxidized to SO_2 and NO_x in the boiler. Included in the list of sulfur and nitrogen reactions products are:

- H_2S
- COS
- CS_2
- NH_3
- Cyanides
- Thiocyanates

These compounds are formed from free radicals liberated from the coal molecule as well as by reactions such as:

$$H_2S + CO_2 \rightleftarrows COS + H_2O$$

$$2H_2S + CO_2 \rightleftarrows CS_2 + 2H_2O$$

Close to 70 different gasifiers have been used commercially in the past or are currently under development (Ref. 2-2) (see Table 2.1-3). Among the important characteristics which distinguish one gasifier from another are:

- Bed type
  - Fixed or supported bed (includes moving bed designs)
  - Fluidized bed
  - Entrained bed

TABLE 2.1-3. POPULATION OF LOW/MEDIUM-BTU GASIFIERS

Gasifier type Gasifier name	Licensor/Developer	Status
Fixed-Bed, Dry Ash		
Lurgi	American Lurgi Corp. (USA)	Present commercial operation
Wellman-Galusha	McDowell Wellman Engr. Co. (USA)	Present commercial operation
Chapman (Wilputte)	Wilputte Corp. (USA)	Present commercial operation
Woodall-Duckham/Gas Integrale	Woodall-Duckham, Ltd. (USA)	Present commercial operation
Riley Morgan	Riley Stoker Corp. (USA)	Present demonstration unit testing; commercially available
Pressurized Wellman-Galusha (MERC)	Morgantown Energy Research Center/ ERDA (USA)	Present development unit testing
Foster Wheeler/Stoic	Foster Wheeler Energy Corp. (USA)	Demonstration unit planned
Kilngas	Allis Chalmers Corp. (USA)	Present development unit testing; commercially available
Kellogg Fixed Bed	M. W. Kellogg Co. (USA)	Present development unit testing
GEGAS	General Electric Research and Development (USA)	Present development unit testing
Consol Fixed Bed	Consolidation Coal Co. (USA)	Present development unit testing
IFE Two Stage	International Furnace Equipment Co., Ltd.	Past commercial operation
Kerpely Producer	Bureau of Mines/ERDA (USA)	Past commercial operation
Marischka	Unknown	Past commercial operation; anthracite or coke only
Pintsch Hillebrand	Unknown (Germany)	Past commercial operation
U.G.I. Blue Water Gas	U.G.I. Corp./DuPont (USA)	Past commercial operation; coke only
Power Gas	Power Gas Co. (USA)	Past commercial operation
Wellman Incandescent	Applied Technology (USA)	Present commercial operation
BCR/Kaiser	Unknown	Past development unit testing
Fixed-Bed, Slagging Ash		
BGC/Lurgi Slagging Gasifier	British Gas Council (GB) Lurgi Mineralöltechnik (W. Germany)	Present development unit testing
GFERC Slagging Gasifier	Grand Forks Energy Research Center/ERDA (USA)	Present development unit testing; lignite only
Luena	Unknown	Past commercial operation; coke only
Thyssen Galocsy	Unknown	Past commercial operation; coke only

(Continued)

TABLE 2.1-3. Continued

Gasifier type Gasifier name	Licensor/Developer	Status
Fluidized-Bed, Dry Ash		
Winkler	Davy Powergas Co. (USA)	Present commercial operation
Hygas	Institute of Gas Technology (USA)	Present development unit testing
Synthane	Pittsburgh Energy Research Center/ ERDA (USA)	Present development unit testing
Hydrane	Pittsburgh Energy Research Center/ ERDA (USA)	Present development unit testing
Cogas	Cogas Development Co. (USA)	Present development unit testing
Exxon	Exxon Corp. (USA)	Present development unit testing
BCR Low-Btu	Bituminous Coal Research (USA)	Present development unit testing
CO_2 Acceptor	Consolidation Coal Co. (USA)	Present development unit testing
Electrofluidic Gasification	Iowa State Univ./ERDA (USA)	Present development unit testing
LR Fluid Bed	Unknown (Germany)	Past commercial operation
HRI Fluidized Bed	Hydrocarbon Research Inc. (USA)	Past development unit testing
BASF-Flesch-Demag	Badische Anilin und Soda Fabrik (West Germany)	Past development unit testing
GECB Marchwood	Unknown	Past development unit testing
Heller	Unknown (Germany)	Past development unit testing
Fluidized-Bed, Agglomerating Ash		
U-Gas	Institute of Gas Technology (USA)	Present development unit testing
Battelle/Carbide	Battelle Memorial Institute (USA)	Present development unit testing
Westinghouse	Westinghouse Electric Corp. (USA)	Present development unit testing
City College of NY Mark 1	Hydrocarbon Research Inc./A.M. Squires (USA)	Present development unit testing
Two-stage Fluidized	British Gas Council (England)	Present development unit testing
ICI Moving Burden	Imperial Chemical Industries, Ltd. (England)	Past development unit testing
Entrained-Bed, Dry Ash		
Garrett Flash Pyrolysis	Garrett Research and Development Co. (USA)	Present development unit testing
Bianchi	Unknown (France)	Past development unit testing; lignite only

(Continued)

TABLE 2.1-3. Continued

Gasifier type Gasifier name	Licensor/Developer	Status
Panindco	Unknown (France)	Past development unit testing; lignite only
USBM Annular Retort	Bureau of Mines/ERDA (USA)	Past development unit testing; lignite only
USBM Electrically Heated	Bureau of Mines/ERDA (USA)	Past development unit testing
Entrained-Bed, Slagging Ash		
Koppers-Totzek	Koppers Co. (USA)	Present commercial operation
Bi-Gas	Bituminous Coal Research, Inc. (USA)	Present development unit testing
Texaco	Texaco Development Corp. (USA)	Present development unit testing
Coalex	Inex Resources, Inc. (USA)	Present development unit testing; commercially available
PAMCO/Foster Wheeler	Pittsburgh and Midway Coal Co./ Foster Wheeler (USA)	Present development unit testing
Combustion Engineering	Combustion Engineering (USA)	Present development unit testing
Brigham Young University	Brigham Young University/ Bituminous Coal Research (USA)	Present development unit testing
Babcock and Wilcox	The Babcock and Wilcox Co. (USA)	Past commercial operation
Ruhrgas Vortex	Ruhrgas A. G. (West Germany)	Past commercial operation
IGT Cyclonizer	Institute of Gas Technology (USA)	Past development unit testing
Inland Steel	Inland Steel Co. (USA)	Past development unit testing
USBM, Morgantown	Morgantown Energy Research Center/ERDA (USA)	Past development unit testing
Great Northern Railway	Great Northern Railway Co. (USA)	Past development unit testing
FRS Cyclone	Unknown (England)	Past development unit testing
Molten Media, Slagging Ash		
Kellogg Molten Salt	M. W. Kellogg Co. (USA)	Present development unit testing
Atgas/Patgas	Applied Technology Corp. (USA)	Present development unit testing
Rockgas	Atomics International (USA)	Present development unit testing
Rummel Single Shaft	Union Rheinische Braun Kohlen Kraftstoff A. G. (West Germany)	Past commercial operation
Sun Gasification	Sun Research and Development Co. (USA)	Past development unit testing
Otto-Rummel Double Shaft	Dr. C. Otto and Co.	Past development unit testing

Source: Ref. 2-2

- Operating conditions
  - Pressure: atmospheric or pressurized
  - Temperature

- Gasification media
  - Reactants: steam, air, oxygen, other additives
  - Coal feed/reactant ratios
  - Mode of reactant introduction

- Coal feeding
  - Mode: continuous or intermittent
  - Mechanism: lock hopper, slurry, screw, etc.
  - Location: top or center of gasifier

- Ash removal
  - Mode: continuous or intermittent
  - Ash condition: dry or slagged (fused)
  - Location: from the gasifier or from the product gas stream

- Energy input for gasification
  - Autothermic: energy supplied by partial combustion of the feed coal in the gasifier
  - Electrothermic: energy supplied by electrical resistance heating
  - Solids circulation/heat transfer: energy supplied by external heating and circulation of additives or inert solids

2.1.1.3 Gas Purification--

With respect to producing a clean fuel, the most important part of a coal gasification system is the gas purification operation. The purpose of this operation is to remove undesirable constituents such as particulates, tars, oils, and acid gases from the raw product gas. The modules needed to satisfy these cleanup requirements include:

- particulate removal,
- gas quenching and cooling, and
- acid gas removal.

A. Particulate Removal--Removal of coal dust, ash and tar aerosols entrained in the raw product gas leaving the gasifier is the primary function of this module. Specific processes commonly used to accomplish this are:

- cyclones,
- electrostatic precipitators (ESP), and
- water or oil scrubbers.

B. Gas Quenching and Cooling--In the gas quenching and cooling module, tars and oils are condensed and particulates and other impurities such as ammonia and cyanides are scrubbed from the raw product gas. Quenching involves the direct contact of the hot raw gas with an aqueous or an organic quench liquor. Extensive cooling of the gas stream occurs initially, primarily through vaporization of the quenching medium. Further gas cooling can be accomplished using waste heat boilers followed by air- and/or water-cooled heat exchangers.

C. Acid Gas Removal--Acid gases such as H_2S, COS, CS_2, mercaptans, and SO_2 are removed from the raw product gas in this module. The commercially available acid gas removal processes, all of which operate at temperatures below 150°C, can be divided into the following categories:

- Physical Solvent Processes
- Chemical Solvent Processes
- Combination Chemical/Physical Solvent Processes
- Direct Conversion Processes
- Catalytic Conversion Processes
- Fixed-Bed Adsorption Processes

Table 2.1-4 lists 70 low-temperature acid gas removal processes and their development status identified from available information (Ref. 2-2). A brief description of the processes in each of the six categories is given in the following text.

Physical solvent processes remove acid gases from the raw product gas by physical absorption in an organic solvent. They are most suited for high pressure applications because the solubilities of acid gases in the solvents are directly proportional to the acid gas partial pressure. Most of the solvents used in these processes have an appreciably higher affinity for H_2S than for CO_2, and can, therefore, be used in a manner that allows for selective removal of H_2S.

Chemical solvent processes remove acid gases by adsorption followed by formation of a chemical complex between the acid gas and solvent. In most of these processes the solvent is regenerated by thermal decomposition of the chemical complex. These processes are generally identified by the type of solvent used. Amine, ammonia, and alkaline salt solutions are the three solvents in common use.

Combination chemical/physical solvent processes use a physical solvent together with an alkanolamine chemical solvent additive. Both the chemical and the physical solvents remove CO_2, H_2S, and HCN; but the physical solvent is primarily responsible for the removal of acid gases such as CS_2, mercaptans, and COS.

Direct conversion processes remove H_2S by adsorption, with subsequent oxidation of the sorbed H_2S to elemental sulfur. These direct conversion processes are divided into two general categories: dry oxidation and liquid phase oxidation.

Catalytic conversion processes are divided into two categories: a) those that convert organic sulfur to H_2S, and b) those that convert organic sulfur and H_2S to SO_2. Most of these processes are generally not considered

TABLE 2.1-4. LOW-TEMPERATURE ACID GAS REMOVAL PROCESSES

Process Category	Process Name and Status
Physical Solvent	Selexol[a,e] Fluor solvent[a] Purisol[a] Rectisol[a,e] Estasolvan[a] Union Oil[b]
Chemical Solvent	
- Amine Solvent	Monoethanolamine (MEA)[a] Diethanolamine (DEA)[a] Triethanolamine (TEA)[a] Methyldiethanolamine (MDEA)[a] Glycol-amine[a] Diisopropanolamine (DIPA)[a] Diglycolamine (DGA)[a]
- Alkaline Salt Solution	Caustic Wash[a] Seaboard[c] Vacuum Carbonate[c] Hot Potassium Carbonate[a] Catacarb[a] Tripotassium Phosphate[c] Benfield[a,e] Alkazid[a] Sodium Phenolate[c] Lucas[a]
- Ammonia Solution	Chemo Trenn[a] Collins[a]
Combination Chemical/Physical Solvent	Amisol[a] Sulfinol[a]
Direct Conversion	
- Dry Oxidation	Iron Oxide (Dry Box)[a] Activated Carbon[a] Claus[a] Great Lakes Carbon Co.[b]
- Liquid Oxidation	Burkheiser[c] Ferrox[c] Knoxo[b] Gludd[c] Manchester[c] Cataban[d] Thylox[c]

TABLE 2.1-4. Continued

Process Category	Process Name and Status
Direct Conversion	
- Liquid Oxidation (Cont'd)	Giammarco-Vetrocoke[a]
	Fischer[a]
	Staatsmijnen-Otto/Autopurification[a]
	Perox[c]
	Stretford[a,e]
	Takahax[a]
	CAS[d]
	Townsend[d]
	Wiewiorowski[d]
	Sulfonly[d]
	Nalco[d]
	Sulphoxide[d]
	Permanganate and Dichromate[a]
	Lacey-Keller[d]
	Sulfox[d]
	Direct Oxidation[a]
Catalytic Conversion	
- Organic Sulfur to H_2S	Carpenter Evans[a]
	Peoples Gas Co.[a]
	Holmes-Maxted[a]
	British Gas Council[d]
	Iron Oxide Catalysts[a]
	Chromia-Aluminum Catalysts[d]
	Copper-Chromium-Vanadium Oxide Catalysts[d]
	Cobalt Molybdenum Catalysts[a]
- Organic Sulfur to H_2S and SO_2	Appleby-Frodingham[a]
	Katasulf[a]
	North Thames Gas Board[a]
	Soda Iron[a]
Fixed-Bed Adsorption	Activated Carbon[a]
	Haines[d]
	Molecular Sieve[a]
	Zinc Oxide[a]

[a] Commercially Available
[b] Under Development
[c] Obsolete/Inactive
[d] Pilot Plant
[e] Have been used or are being used in commercial or demonstration gasification plants

Source: Ref. 2-2, 2-17, 2-19 through 2-35

to be acid gas removal processes; however, they can be used to convert hard-to-remove acid gases such as COS, CS_2, and mercaptans into compounds such as H_2S and SO_2, which can then be handled by other acid gas removal processes.

Fixed-bed adsorption processes remove acid gases by adsorption on a fixed sorbent bed. The amount of acid gases removed is dependent on the surface area available for adsorption. Regeneration of the sorbent is generally accomplished by thermal methods.

2.1.1.4 Pollution Control Operations--

Air emissions, liquid effluents, and solid wastes from the process operations just described will require pollution control modules. The function of these modules is to achieve levels of control that are consistent with environmentally acceptable plant practices.

A. Air Pollution Control--Air pollutants from low-Btu gasification processes are primarily coal dust, coal feeder vent gases, combustion gases, process tail gases and tank vents. These streams are processed in various combinations of control modules to achieve particulate control, sulfur control and recovery, hydrocarbon control and nitrogen oxides control. A flow diagram of these modules is presented in Figure 2.1-3. In this figure, the gaseous effluents which may be directed to these four modules and potential flow paths between the modules are identified. The nature of the contaminated gaseous effluents dictate which modules are required to treat the gases.

1) Particulate Control Module - Coal dust from the coal pretreatment and coal gasification operations are the principal particulate emissions requiring control. Other emission sources include the ash handling system and the permanent coal storage pile. The severity of the particulate emission problem will vary from site to site. Water sprays are used at coal conveying transfer points at some sites; however, these may or may not be effective control devices.

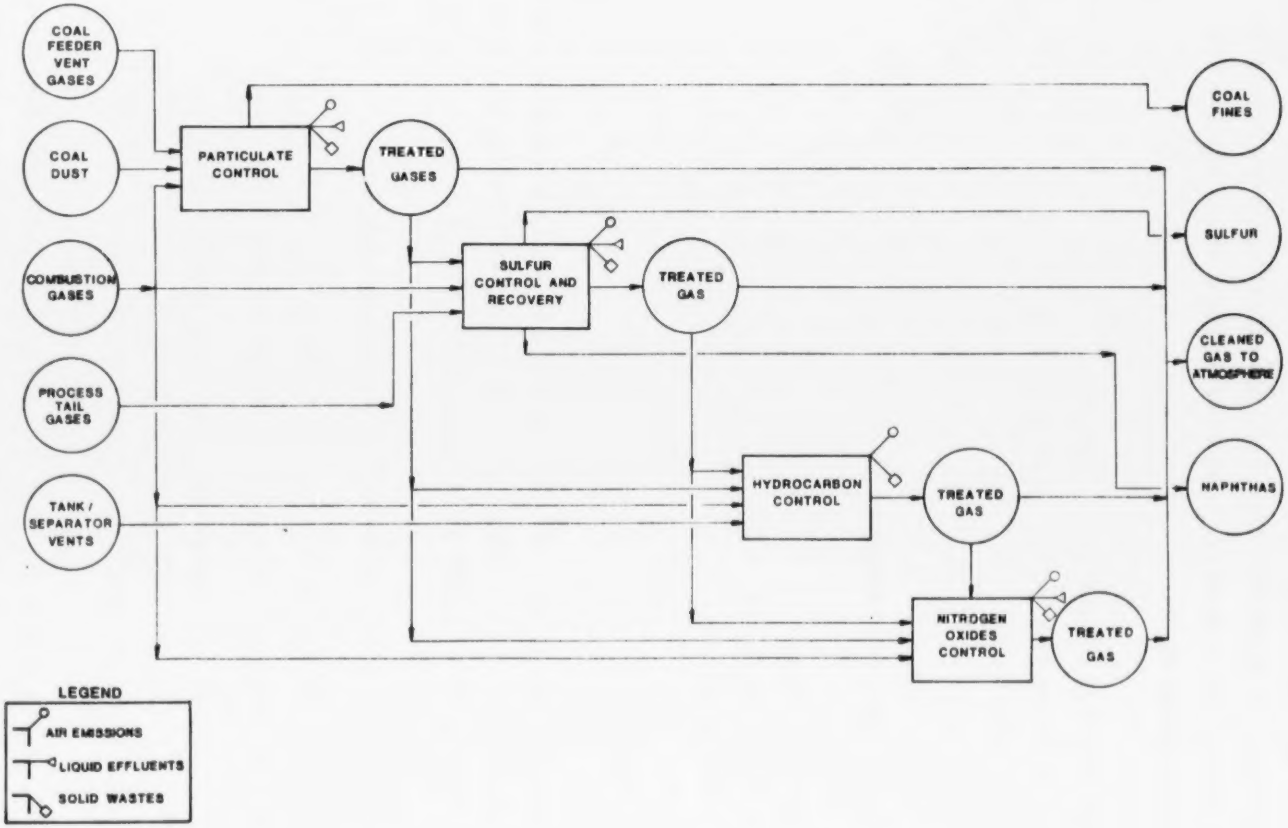

COAL FEEDER VENT GASES
COAL DUST
COMBUSTION GASES
PROCESS TAIL GASES
TANK / SEPARATOR VENTS
PARTICULATE CONTROL
TREATED GASES
SULFUR CONTROL AND RECOVERY
TREATED GAS
HYDROCARBON CONTROL
TREATED GAS
NITROGEN OXIDES CONTROL
TREATED GAS
COAL FINES
SULFUR
CLEANED GAS TO ATMOSPHERE
NAPHTHAS
LEGEND
AIR EMISSIONS
LIQUID EFFLUENTS
SOLID WASTES

Figure 2.1-3. Flow diagram for the modules in the air pollution control operation.

The many processes and variations of processes that could be used to control particulate emissions from coal gasification processes are generally divided into the following four categories, based on the collection mechanism used:

- Mechanical collectors
  - Settling chambers
  - Cyclones
  - Filters
- Electrostatic precipitators
- Wet collectors
  - Spray chambers
  - Wet scrubbers
- Afterburners

2) Sulfur Control Module - All operations in low- or medium-Btu coal gasification plants are potential sources of sulfur-bearing gaseous effluents. Examples of these effluents are:

- Tail gases from the acid gas removal module,
- On-site power generation flue gases,
- Vent gases from the water pollution control module,
- Coal feeder vent gases from the coal gasification module, and
- Gases from the particulate control module.

The function of the sulfur control module is to reduce the concentrations of the sulfur compounds such as H_2S, COS, CS_2, and SO_2 to levels acceptable for discharge to the environment.

The processes capable of removing sulfur compounds from gas streams can be divided into three general categories.

- primary sulfur recovery processes,
- tail gas cleanup processes (secondary recovery), and
- sulfur oxides control processes.

The principles of operations of the sulfur control processes are discussed in the following paragraphs.

There are numerous processes based on removal of sulfur compounds from gas streams, followed by recovery of the sulfur as a by-product. These direct conversion processes can be classified as either dry oxidation or liquid phase oxidation and are listed in Table 2.1-5. The principle of operation involves the oxidation of sulfur compounds to elemental sulfur, which is a salable by-product. The two most widely used direct conversion processes are the Claus (dry oxidation) and the Stretford (liquid phase oxidation) processes.

TABLE 2.1-5. DIRECT CONVERSION PRIMARY SULFUR RECOVERY PROCESSES

Dry Oxidation Processes

Iron Oxide (Dry Box)
Activated Carbon
Claus
Sulfreen
Great Lakes Carbon

Liquid Oxidation Processes

Burkheiser
Ferrox
Konox
Fludd
Manchester
Cataban
Thylox
Giammarco-Vetrocoke
Fischer
Staatsmijnen-Otto
Autopurification
Perox
Stretford
Takahax
C.A.S.
Townsend
Wiewiorowski
Sulfonly
Nalco
Sulphoxide
Permanganate and Dichromate
Lacey-Keller
Sulfox
Direct Oxidation

References: 2-19, 2-23, 2-29

Tail gas cleanup processes are used to remove and, in some cases, recover the sulfur compounds remaining in the tail gases of primary sulfur recovery processes. These processes, when combined with a Claus unit for example, can provide an overall sulfur removal effectiveness of up to 99.9+ percent. Commercially available tail gas cleanup processes are classified as follows:

Process Type	Process Name
Removal of sulfur compounds and recovery of elemental sulfur	Beavon Cleanair CBA Sulfreen
Reduction of sulfur compounds to H_2S which is recycled to a Claus unit	SCOT Trencor-M

Sulfur oxides control processes are not major functions within coal gasification plants. They are primarily flue gas desulfurization processes and are generally used to control sulfur emissions from on-site coal-fired heaters and boilers.

3) Hydrocarbon Control Module - The function of this module is to reduce the hydrocarbon content of process tail gases, vent streams and other waste streams to levels acceptable for discharge to the environment. There are two basic methods of hydrocarbon control: afterburners and adsorbers. Afterburners simply convert hydrocarbons to CO_2 and H_2O by oxidation. Adsorbers use sorbents such as activated carbon to remove the hydrocarbons from the gas stream.

4) Nitrogen Oxides Control Module - A nitrogen oxide control strategy for the combustion gases emitted from coal or low/medium-Btu gas-fired boilers and furnaces may be required. NO_x formation in the gasification module is expected to be low since the raw gas passes through a reducing atmosphere before leaving the gasifier. The nitrogen that does react in

the gasifier should form NH_3, HCN, thiocyanates, and other nitrogen-containing organics rather than nitrogen oxides (Ref. 2-36). There are three basic processes that can be used to control NO_x emissions from boilers and furnaces:

- Combustion modifications,
- Post-combustion flue gas cleaning, and
- Fluidized-bed combustion.

These are processes which would not be considered central to those in coal gasification plants.

B. Water Pollution Control--In a coal gasification facility, the specific sources which generate wastewaters will determine the type of contaminants that are present in those streams. Wastewater sources in a coal gasification plant are shown in Figure 2.1-4 along with descriptions of the particular type of wastewaters they generate. Typical wastewaters and the modules required to treat them are shown schematically in Figure 2.1-5.

The suspended solid contaminants are primarily particulates that are generated when the coal is crushed and sized and/or when ash is quenched as it is discharged from the gasifier. Dissolved and suspended organics are volatile hydrocarbons that are condensed in the quench liquor during the subsequent raw gas cooling step. Dissolved inorganic gas contaminants such as CO_2, H_2S, and NH_3 are produced in the same manner as the dissolved organics. Dissolved salts accumulate when reuse of the upgraded wastewaters is maximized. At higher concentrations, salts begin to scale-out on heat exchanger and process equipment surfaces; consequently, close monitoring of dissolved solids in the wastewater will be an essential control practice.

The various types of processes which can be used to satisfy the requirements of the four water pollution control modules are listed below:

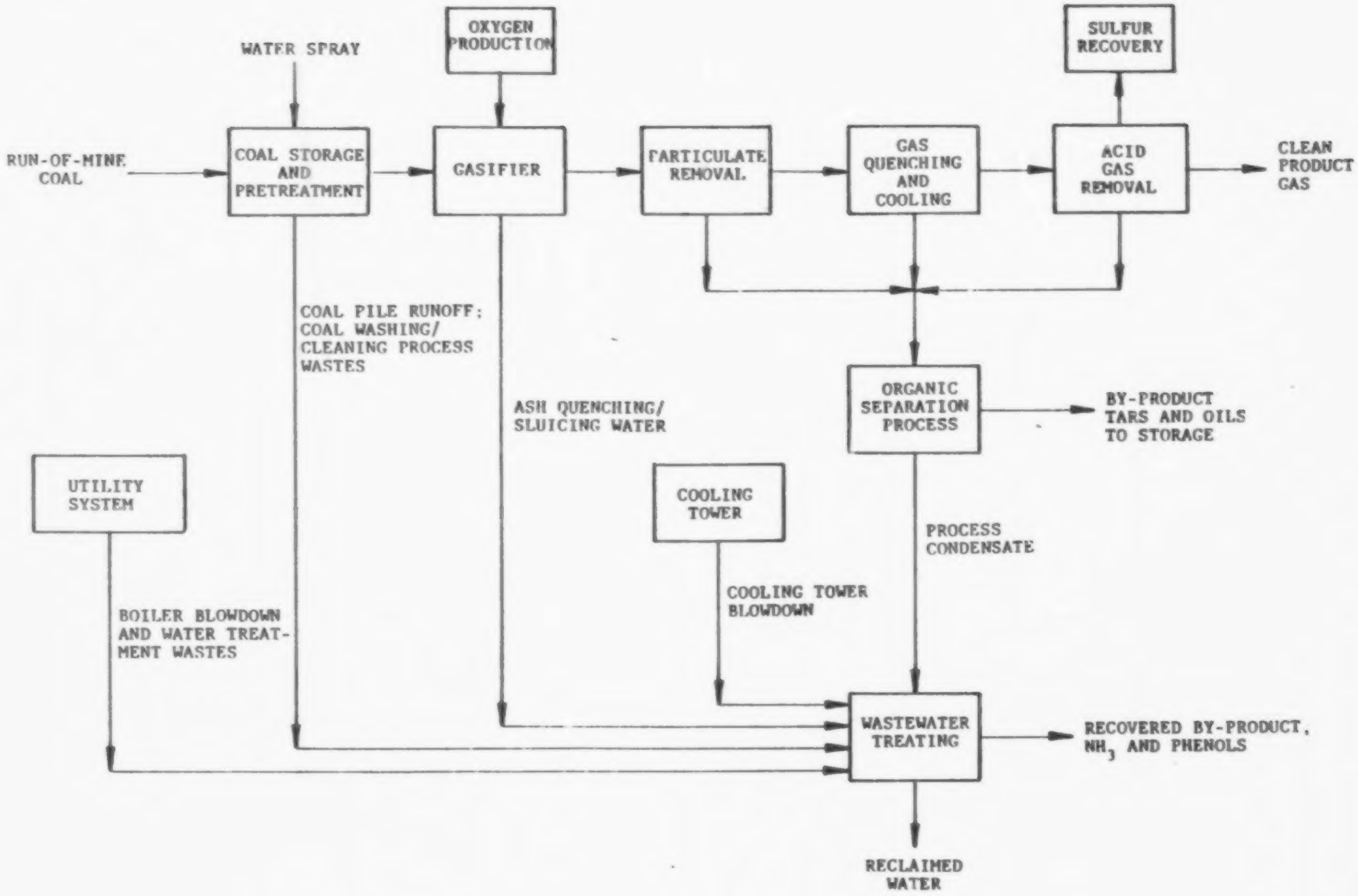

OXYGEN PRODUCTION
SULFUR RECOVERY
WATER SPRAY
RUN-OF-MINE COAL
COAL STORAGE AND PRETREATMENT
GASIFIER
PARTICULATE REMOVAL
GAS QUENCHING AND COOLING
ACID GAS REMOVAL
CLEAN PRODUCT GAS
COAL PILE RUNOFF; COAL WASHING/ CLEANING PROCESS WASTES
ORGANIC SEPARATION PROCESS
BY-PRODUCT TARS AND OILS TO STORAGE
ASH QUENCHING/ SLUICING WATER
UTILITY SYSTEM
COOLING TOWER
PROCESS CONDENSATE
COOLING TOWER BLOWDOWN
BOILER BLOWDOWN AND WATER TREATMENT WASTES
WASTEWATER TREATING
RECOVERED BY-PRODUCT, NH_3 AND PHENOLS
RECLAIMED WATER

Figure 2.1-4. Major process modules generating wastewater in a typical coal gasification plant.

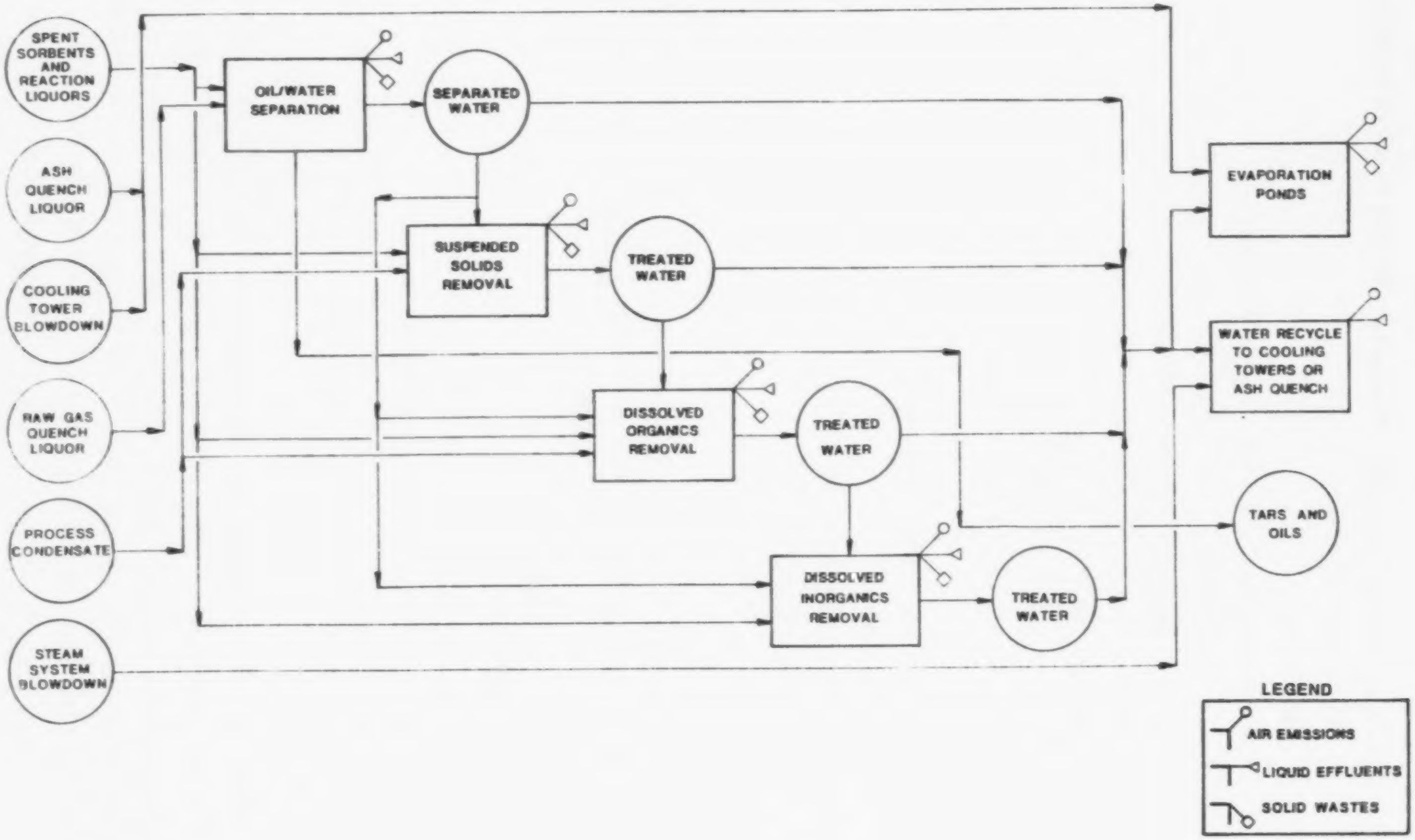

SPENT SORBENTS AND REACTION LIQUORS
ASH QUENCH LIQUOR
COOLING TOWER BLOWDOWN
RAW GAS QUENCH LIQUOR
PROCESS CONDENSATE
STEAM SYSTEM BLOWDOWN
OIL/WATER SEPARATION
SEPARATED WATER
SUSPENDED SOLIDS REMOVAL
TREATED WATER
DISSOLVED ORGANICS REMOVAL
TREATED WATER
DISSOLVED INORGANICS REMOVAL
TREATED WATER
EVAPORATION PONDS
WATER RECYCLE TO COOLING TOWERS OR ASH QUENCH
TARS AND OILS
LEGEND
AIR EMISSIONS
LIQUID EFFLUENTS
SOLID WASTES

Figure 2.1-5. Flow diagram for the modules in the water pollution control operation.

- Oil/Water Separation and Suspended Solids Removal Modules
  - Oil/water separator
  - Filtration
  - Flocculation/flotation (dissolved air)

- Dissolved Organics Removal Module
  - Extraction
  - Adsorption
  - Biological treatment
  - Cooling tower oxidation (stripping)

- Dissolved Inorganics Removal Module
  - Stripping
  - Brine concentration
  - Ion exchange
  - Membrane desalination

C. Solid Waste Pollution Control--The solid waste pollution control module treats and disposes of the following classes of wastes:

- Ash
- Coal residue
- Biological oxidation sludge
- Spent catalysts and filter media
- Coal fines
- Sulfur

Coal fines may be collected and burned on site; coal fines and sulfur may be sold as by-products. The other wastes may or may not require treatment before disposal, depending upon their composition. Chemical fixation and sludge reduction modules can be used to treat these solid wastes. Figure 2.1-6 is a flow diagram of the modules for solid waste control. Landfill, by definition, is the ultimate disposal technique for these wastes.

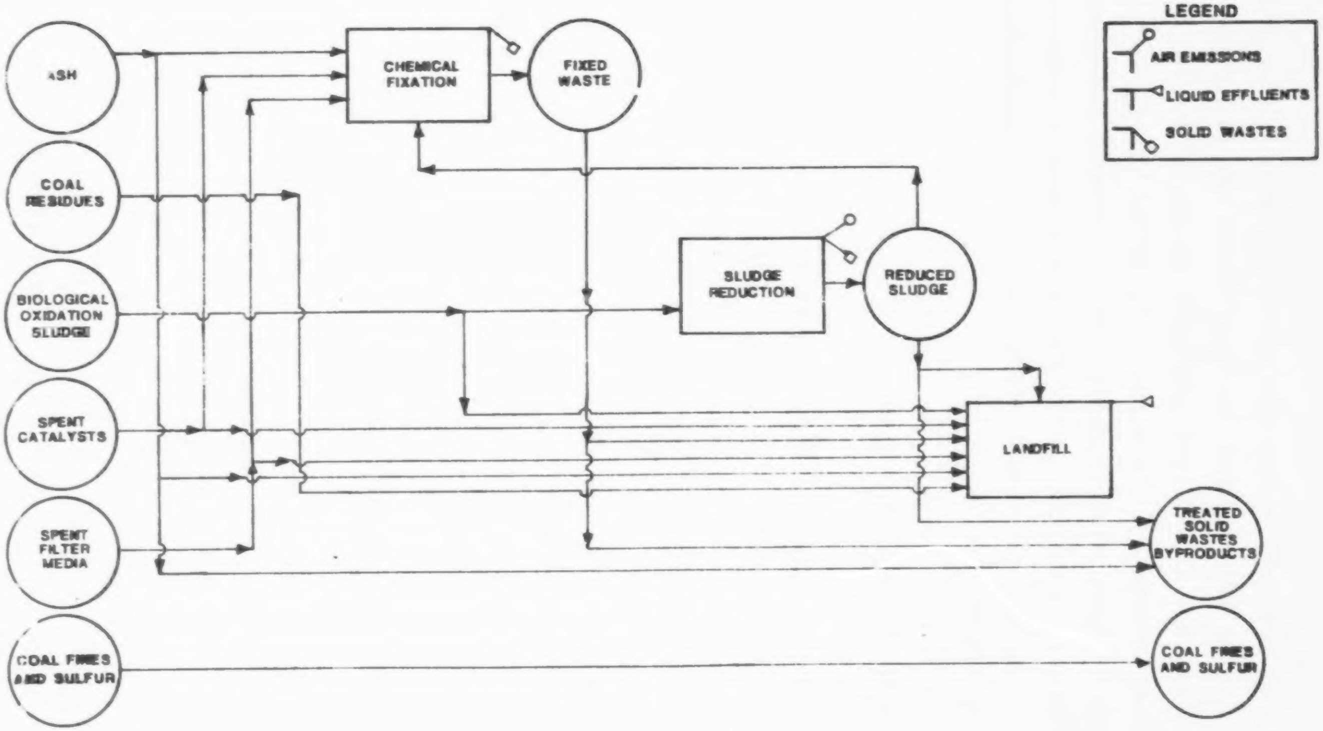

ASH
COAL RESIDUES
BIOLOGICAL OXIDATION SLUDGE
SPENT CATALYSTS
SPENT FILTER MEDIA
COAL FINES AND SULFUR
CHEMICAL FIXATION
FIXED WASTE
SLUDGE REDUCTION
REDUCED SLUDGE
LANDFILL
TREATED SOLID WASTES BYPRODUCTS
COAL FINES AND SULFUR
LEGEND
AIR EMISSIONS
LIQUID EFFLUENTS
SOLID WASTES

Figure 2.1-6. Flow diagram for the modules in the solid waste control operation.

2.1.2 Status of Technology

The production of low- and medium-Btu gas from coal has been practiced both in the United States and overseas for many years. It is estimated that at one time there were some 11,000 coal gasifiers in use in the U.S. But as the availability of natural gas increased, the number of operating gasification systems declined significantly. At the present time there are only a few coal gasifiers operating in the United States on a commercial basis (Ref. 2-37). However, in general low/medium-Btu coal gasification can be regarded as being commercially available.

2.1.2.1 Status of Coal Pretreatment Processes--

The coal pretreatment processes which may be required in a low- or medium-Btu coal gasification plant (see Table 2.1-1), with the exception of partial oxidation, have been used for years in the gasification, coal-fired electric utility and charcoal industries (Ref. 2-38). However, much work, e.g., the Bureau of Mines work with the Synthane Process, has been done to develop a reliable partial oxidation process for reducing the caking tendencies of certain types of coal (Ref. 2-39). Therefore, the development status of all of the coal pretreatment processes is such that they can be regarded as commercially available.

2.1.2.2 Status of Gasification Processes--

Low- and medium-Btu gasification systems being considered today are in varying stages of development. The development status of the 70 gasifiers which have been used commercially in the past or are currently under development are shown in Table 2.1-3 (Ref. 2-2). A number of the smaller systems are commercially available, and a few (Wellman-Galusha and Chapman) are operating in the United States. Several of the other systems, e.g. the Lurgi and the Koppers-Totzek, have been or are currently operating commercially in foreign countries.

As part of an environmental assessment of low- and medium-Btu gasification technology, Radian Corporation has compiled a list of fourteen gasifiers which appear to be the most promising candidates for satisfying near-term commercial needs for low/medium-Btu gas. This list is shown in Table 2.1-6.

TABLE 2.1-6. PROMISING LOW- AND MEDIUM-BTU GASIFICATION SYSTEMS

First Group[1]	Second Group[2]	Third Group[3]
Wellman-Galusha	Chapman (Wilputte)	Pressurized Wellman-Galusha (MERC)
Lurgi	Riley Morgan	
Woodall Duckham/ Gas Integrale		BGC/Lurgi Slagging Gasifier
Koppers-Totzek		Texaco
Winkler		Bi-Gas
Wellman Incandescent		Coalex
Foster-Wheeler/Stoic		

[1] Commercially available; significant number of units currently operating in the U.S. or in foreign countries.

[2] Commercially demonstrated in limited applications.

[3] Commercial or demonstration-scale units operating or being constructed; technology is promising and should be monitored.

Source: Ref. 2-37

The major criterion used in making these selections was development status (Ref. 2-2). There are many gasifiers which are not likely to be used in the near future. These include gasifiers that have been operated in the past but have been abandoned for various reasons. Other gasifiers are in such early stages of development that their technical and economic feasibility cannot be adequately assessed. All gasifiers that are not either commercially available or currently being tested on a demonstration or large pilot-plant scale were not included in the list of promising processes (Ref. 2-2). While the major factor in Radian's selections was development status, consideration was also given to 1) applicability to low/medium-Btu gasification, 2) energy efficiency, 3) process limitations, 4) environmental impacts and 5) costs (Ref. 2-2).

2.1.2.3 Status of Gas Purification Processes--

The processes for quenching/cooling and removing particulates from raw low-or medium-Btu gas are commercially proven (Ref. 2-40). For acid gas removal, either high (>150°) or low temperature processes can be used in theory. However, while high temperature processes have thermal advantages, they are unfortunately only in the development stage. For this reason existing and proposed gasification plants use low temperature processes (Ref. 2-5, 2-41, 2-42, 2-43, 2-44). Although most of the available processes (see Table 2.1-4) were developed for use in the natural gas, coking, refinery and chemical process industries, there do not appear to be any prohibitive technical problems associated with their use in coal gasification plants.

In a previous Radian environmental assessment of low- and medium-Btu gasification technology, fourteen promising acid gas removal processes were selected from the low temperature acid gas removal systems summarized in Table 2.1-4. These fourteen processes are listed in Table 2.1-7.

TABLE 2.1-7. PROMISING ACID GAS REMOVAL SYSTEMS

Process category	Process name
Physical solvent	Rectisol Selexol Purisol Estasolvan Flour solvent
Chemical solvent	MEA (monoethanolamine) MDEA (methyldiethanolamine) DEA (diethanolamine) DIPA (diisopropanolamine) DGA (diglycolamine) Benfield
Combination chemical/physical solvent	Amisol Sulfinol
Direct conversion	Stretford

Source: Ref. 2-2

The criteria for the selection of these fourteen systems were their applicability to low/medium-Btu gasification, development status, environmental impacts, energy requirements, costs, and process limitations (Ref. 2-2).

2.1.3 Applicability and Limitations

Both low- and medium-Btu gas are suitable fuel sources for industrial boilers (Ref. 2-45). While their combustion characteristics are somewhat different than those of natural gas, those differences can be compensated for in the design of a new boiler. For retrofit situations use of either synthetic fuel will require at least modification or replacement of the burners and fuel lines. Use of low-Btu gas will also slightly derate (<10% for a 6 MJ/m^3 low-Btu gas) a natural gas-fired boiler (Ref. 2-46). However, these disadvantages are off-set by the capital cost savings that result from being able to use the existing boiler.

There are several important factors to be considered in making a choice between using low- and medium-Btu gas. First, and most important, is the cost of the fuel gas. Industrial boiler operators are cost sensitive and are likely to select the system which offers the lowest cost. Other factors which are also important include the availability of land and environmental considerations.

Medium-Btu gasification systems require a greater capital investment per unit of energy produced than low-Btu systems. This is due mainly to the medium-Btu system's need for an oxygen production unit, but also reflects the greater complexity of medium-Btu systems. In order to take advantage of economies of scale and hence reduce the cost of the product gas, medium-Btu gasification facilities tend to be large capacity installations. Such is the case for the Lurgi medium-Btu plants currently operating overseas in Yugoslavia and South Africa. In comparison to the fuel needs of even a large industrial boiler (117 MW_T heat input or 400 x 10^6 Btu/hr), a medium-Btu gasification facility would have a much larger capacity ($\geq$1500 MW_T or 5.1 x 10^9 Btu/hr). On the other hand, low-Btu systems typically have capacities similar to those of industrial boilers: 10 to 118 MW_T ($30-400 \times 10^6$ Btu/hr). Thus, unless the medium-Btu gas could be used in more than one industrial boiler, low-Btu gas appears to be a better choice.

In addition to costs, land availability and environmental considerations are important factors. Although a low-Btu gasification facility does not require substantial land area, a gasification/boiler facility will require more land than a boiler by itself. If this additional land is not available, an alternative would be to site the fuel generation facility offsite and pipe the fuel gas to the boiler. Also, there are environmental discharges associated with a gasification facility. Depending on local conditions, it may be impossible to install a new emission source at or near the boiler site. If so, off-site fuel generation would again be an alternative. In either of these situations where the fuel gas must be transported, medium-Btu gas has an economic advantage over low-Btu gas because of its greater energy content ($\sim$13 MJ/m^3 vs $\sim$6 MJ/m^3 for low-Btu gas), and hence lower transportation costs.

2.1.4 Factors Affecting Performance

The performance of low- and medium-Btu coal gasification systems as emission control techniques for industrial boilers depends on the performance of the gas purification operation. And, more specifically, it depends on the performance of the acid gas removal unit in removing H_2S and organic sulfur compounds (predominantly COS) from the product gas. Under proper operating conditions, *i.e.*, temperatures, pressures, etc., the chemical, physical, and combination chemical/physical solvent acid gas removal processes listed in Table 2.1-7 are capable of effecting essentially complete removal of sulfur species from the synthetic gases (Ref. 2-2). Unfortunately, the minimum operating pressure for complete sulfur species removal for these processes ranges from around 1.4 MPa for some of the amine processes to around 6.9 MPa for most of the physical solvent processes (Ref. 2-23). While at operating pressures below those just stated, the system's performance is lessened, the extent to which system performance is affected is not known. However, because the amine processes use chemical means to absorb acid gases, they are affected less by pressure than the physical solvent processes.

The removal efficiency of direct conversion processes, such as the Stretford process, is not significantly affected by the gas stream pressure (Ref. 2-23). However, while these processes are very effective in removing H_2S, they are not effective in removing organic sulfur compounds (Ref. 2-2). Thus, the performance of a gasification system employing a direct conversion acid gas removal process is limited by the organic sulfur content of the raw product gas.

2.1.5 System Performance

Currently, there are no data available on the actual emissions resulting from combustion of low- or medium-Btu fuel gas derived from coal in industrial boilers. However, based on typical fuel gas compositions and limited small-scale tests, estimates of emissions can be made. These estimates as well as a discussion of the impact of utilizing low- and medium-Btu gas in industrial boilers are presented in the following sections. For all situations addressed here, it is assumed that the gasification system includes quenching/cooling and acid gas removal.

2.1.5.1 Emissions Reduction Potential--

By incorporating particulate removal and quenching/cooling in a coal gasification system, essentially all of the entrained particulates, and 99+% of the tars and oils can be removed from the low- or medium-Btu product gas. This can be accomplished by using cyclones (60-80 percent removal of the incoming particulates and heavy tars), water or oil scrubbers (60-70 percent removal of the incoming tars and oils), and/or electrostatic precipitators that can be designed to remove up to 99 percent of the residual particulates, tars, and oils (Ref. 2-36, 2-47, 2-48, 2-49). The cleaned synthetic gases are essentially particulate free, and any residual tar and oil aerosols should be destroyed during combustion. Thus, negligible particulate emissions will result. This implies that low- or medium-Btu gasification effects essentially complete control of particulate emissions from the utilization of coal in an industrial boiler.

When coal is combusted, NO_x emissions range from 270 ng NO_x/J (0.63 lb/10^6 Btu) for small boilers to 400 ng/J (0.93 lb/10^6 Btu) for larger industrial boilers (Ref. 2-1). Limited laboratory tests on the combustion of low- or medium-Btu gas have indicated that NO_x emissions are similar to (Ref. 2-36) or less than (Ref. 2-50) those produced by the combustion of natural gas (50-100 ng NO_x/J) (Ref. 2-1). These results are for fuel gases containing very little, if any, ammonia or cyanide compounds. For gasification systems with quenching/cooling and acid gas removal processes, this is a reasonable assumption. However, if ammonia and cyanides are present, they will be oxidized, by varying degrees, to NO_x (Ref. 2-36, 2-50).

When coal is converted into low- or medium-Btu gas, approximately 90-98 percent of the coal sulfur is converted to hydrogen sulfide (H_2S) (Ref. 2-5, 2-6, 2-49). Under the proper operating conditions, the acid gas removal processes listed in Table 2.1-7 can remove essentially all (99+ percent) of the H_2S in a gaseous stream (Ref. 2-2, 2-4, 2-23, 2-51, 2-52, 2-53, 2-54). Thus, low- and medium-Btu coal gasification can effect at least 90+ percent control of sulfur emissions resulting from utilizing coal in industrial boilers.

In terms of SO_2 emissions per unit heat input to the boiler, a representative range for coal derived gases is 40-130 ng SO_2/J (0.09-0.30 lb/10^6 Btu), as determined from residual sulfur levels remaining after treatment by an acid gas removal process. This compares to uncontrolled emissions of 510-2420 ng SO_2/J (1.18-5.62 lb/10^6 Btu) for direct combustion of 0.6-3.5 percent sulfur coal (Ref. 2-1).

2.1.5.2 Impacts on the Boiler--

The use of coal derived low- or medium-Btu gas in a new industrial boiler has several advantages over a direct coal-fired boiler. First, a gas-fired boiler is a much simpler piece of equipment to operate than a coal-fired boiler. There is no need for ash handling equipment and the only fuel handling equipment required is piping. Second, due to the less complex nature of gas-fired boilers, maintenance requirements will be less than for a coal-fired boiler.

On the other hand, the production and use of coal derived gases in an industrial boiler can have adverse impacts. The primary concerns are the reliability/operability of the gasification system and how that affects the desired operability of the boiler. In order to minimize adverse impacts, installation of spare capacity or sparing of key process units in the gasification system may be required. Another alternative would be to provide a backup fuel source (such as distillate fuel oil) for the boiler. Incorporating any of these options into a gasification/steam generation system design must be made on a case by case basis, taking into consideration the particular requirements of the boiler system. In addition, in the selection and design of the boiler, consideration must be given to the different combustion characteristics (e.g., heat release rate and flame temperatures) of coal-derived gases versus natural gas.

2.2 HIGH-BTU COAL GASIFICATION

High-Btu coal gasification is a technology being developed to convert our country's vast coal resources into synthetic natural gas (SNG). High-Btu gasification is merely an extension of medium-Btu gasification. While the latter process produces a product gas with a high CO and H_2 content, high-Btu systems go one step further and convert the H_2 and CO into methane. Since this methanation step (as well as the shift conversion reaction required prior to methanation) involves an exothermic reaction, the energy conversion efficiency of a high-Btu system is lower than a medium-Btu system.

The lower energy conversion efficiency, along with the added capital costs for the methanation and shift conversion units, result in higher costs for high-Btu gas (versus medium-Btu gas). Thus, unless an unusual set of site specific conditions exist, medium-Btu gas is economically more attractive than high-Btu gas for use in an industrial boiler.

In light of the above considerations, it does not seem appropriate to consider high-Btu coal gasification as an emission control technique for industrial boilers. Accordingly, high-Btu gasification is not addressed further in this technology assessment report.

2.3 COAL LIQUEFACTION

Coal liquefaction is currently under development as a segment of the National Energy Plan (NEP). Liquefaction is intended in part to convert fossil fuels such as coal and oil shale to a form compatible with present end uses. It is also intended to permit environmentally acceptable use of presently accessible fossil-fuel supplies which fail to meet environmental requirements in conventional applications (Ref. 2-55). Therefore, in addition to converting forms (solids to liquids) for end-use considerations, the liquefaction facilities can be considered a fuel pretreatment for pollution control purposes.

2.3.1 System Description

Coal liquefaction technology consists of four basic operations and auxiliary processes. These operations are 1) coal pretreatment, 2) coal liquefaction, 3) separation, and 4) purification/upgrading. The processes in these operations are shown in Figure 2.3-1.

2.3.1.1 Coal Pretreatment--

Coal pretreatment processes used in coal liquefaction technology are similar to those required for gasification technology. Coal pretreatment

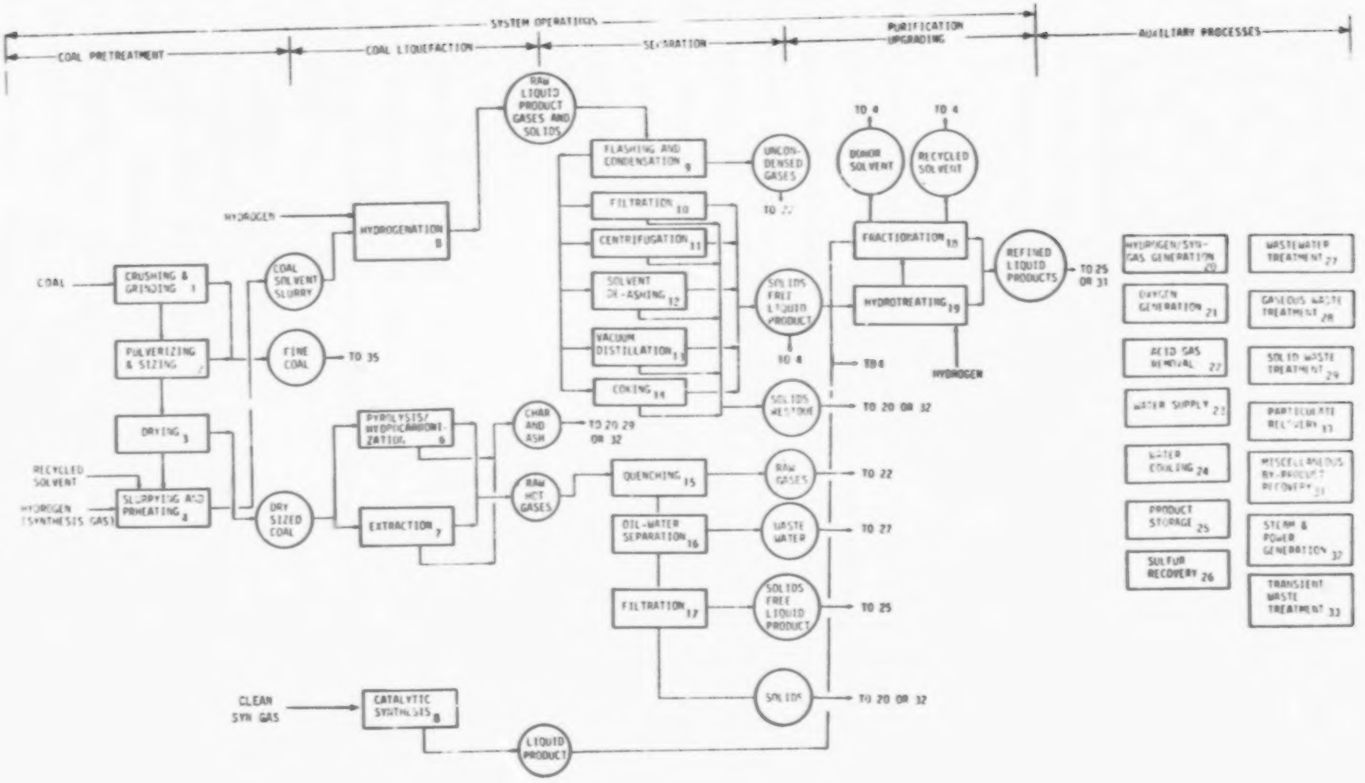

SYSTEM OPERATIONS
COAL PRETREATMENT
COAL LIQUEFACTION
SEPARATION
PURIFICATION UPGRADING
AUXILIARY PROCESSES
COAL
CRUSHING & GRINDING 1
PULVERIZING & SIZING 2
DRYING 3
RECYCLED SOLVENT
HYDROGEN (SYNTHESIS GAS)
SLURRYING AND PRHEATING 4
COAL SOLVENT SLURRY
FINE COAL
TO 35
DRY SIZED COAL
HYDROGEN
HYDROGENATION 8
PYROLYSIS/ HYDROCARBONI- ZATION 6
EXTRACTION 7
CLEAN SYN GAS
CATALYTIC SYNTHESIS 8
LIQUID PRODUCT
RAW LIQUID PRODUCT GASES AND SOLIDS
CHAR AND ASH
TO 20 29 OR 32
RAW HOT GASES
FLASHING AND CONDENSATION 9
FILTRATION 10
CENTRIFUGATION 11
SOLVENT DE-ASHING 12
VACUUM DISTILLATION 13
COKING 14
QUENCHING 15
OIL-WATER SEPARATION 16
FILTRATION 17
UNCON- DENSED GASES
TO 22
SOLIDS FREE LIQUID PRODUCT
TO 4
SOLIDS RESIDUE
TO 20 OR 32
RAW GASES
TO 22
WASTE WATER
TO 27
SOLIDS FREE LIQUID PRODUCT
TO 25
SOLIDS
TO 20 OR 32
TO 4
TO 4
TO 4
DONOR SOLVENT
RECYCLED SOLVENT
FRACTIONATION 18
HYDROTREATING 19
HYDROGEN
REFINED LIQUID PRODUCTS
TO 25 OR 31
HYDROGEN/SYN- GAS GENERATION 20
OXYGEN GENERATION 21
ACID GAS REMOVAL 22
WATER SUPPLY 23
WATER COOLING 24
PRODUCT STORAGE 25
SULFUR RECOVERY 26
WASTEWATER TREATMENT 27
GASEOUS WASTE TREATMENT 28
SOLID WASTE TREATMENT 29
PARTICULATE RECOVERY 30
MISCELLANEOUS BY-PRODUCT RECOVERY 31
STEAM & POWER GENERATION 32
TRANSIENT WASTE TREATMENT 33

Figure 2.3-1. Module diagram for coal liquefaction systems.

can include crushing/grinding, pulverizing, sizing, drying, and/or slurrying/preheating.

2.3.1.2 Coal Liquefaction--

There are four basic processes that can be used in the coal liquefaction operation. These processes are: hydrogenation, pyrolysis/hydrocarbonization, extraction, and catalytic synthesis, which is gasification followed by synthesis and other processes.

A. Hydrogenation--Hydrogenation is gaining prominence over other systems because it is in an advanced stage of development, provides more flexible utilization of product fuel, has higher overall conversion efficiencies, and shows better market potential and economic advantages. Hydrogenation involves direct conversion of coal to liquids through the addition of hydrogen to coal at elevated temperature and pressure. The three types of hydrogenation systems, which are distinguished by the hydrogen transfer method used in the reactor, are: noncatalytic, catalytic, and donor solvent.

In noncatalytic hydrogenation, coal is partially dissolved in a hydrogen-rich solvent and the undissolved solids are filtered out. The solvent is recovered from the high-boiling product and recycled to the dissolver. In a commercial plant the carbonaceous solids will be gasified to produce hydrogen. The Solvent Refined Coal (SRC) process uses this method. Two types of SRC processes are being developed: SRC-I and SRC-II. In SRC-I, slurried coal is liquified and the product is separated from the unreacted residue by filtration. Recycled solvent for coal-slurry preparation is recovered from the product mixture by distillation. The rest of the liquid product is solidified to produce a boiler fuel having lower sulfur and ash content than the base coal. In SRC-II, more hydrogen is added to coal in the liquefaction reactor. The process consumes almost twice the amount of hydrogen that is consumed in the SRC-I process (Ref. 2-56). The unreacted solids are separated from the product by vacuum distillation. A fraction of the liquid product is recycled for slurry preparation. The product liquids may be hydroprocessed for further upgrading, depending upon the product quality desired.

In catalytic hydrogenation, pulverized coal is slurried with coal-derived recycled oil, mixed with hydrogen, and fed to a catalytic reactor operating at moderate temperatures and high pressures. The recycled aromatic solvent recovered from the process does not require hydrotreating. The H-Coal and Synthoil processes use catalytic hydrogenation.

In the third type of hydrogenation, the donor solvent system, no catalyst and only a small amount of molecular hydrogen are used in the reactor. The feed coal reacts with the donor hydrogen from the solvent and yields liquid products. The solids-free liquid product from the reactor is hydrotreated in the presence of a catalyst, and part of this hydrotreated liquid product is used as recycled slurry solvent. The Exxon Donor Solvent (EDS) and Consol Synthetic Fuels (CSF) processes employ this method.

B. Carbonization--In carbonization systems, coal is pyrolyzed to recover products by application of heat, either without direct addition of hydrogen (pyrolysis), or with the addition of hydrogen (hydrocarbonization). In this system, most of the carbon remains as a solid. Liquids and gases containing a higher hydrogen-to-carbon (H/C) ratio than the original coal are recovered. Pyrolysis processes produce significant quantities of by-product gas and char. The liquid product is further hydrogenated for desulfurization and quality improvement. Char-Oil Energy Development (COED), Clean Coke, and Occidental Research Corporation (ORC) flash pyrolysis processes use this method of coal liquefaction.

C. Extraction--In the Supercritical Gas Extraction Process, extractant gas near its supercritical temperature is used to extract liquid from coal. Advantages of this system include easy separation (by cooling and depressurization) of the liquid products and extractant gas, and production of products which contain more hydrogen and are of lower molecular weight than the products of other processes. Unfortunately, the SGEP has not been developed beyond the laboratory stage.

D. Catalytic Synthesis--In the system which uses gasification of coal followed by syngas synthesis, carbon monoxide and hydrogen react in the presence of the Fischer-Tropsch catalyst to produce a wide variety of liquid products such as LPG (C_3-C_4), gasoline (C_5-C_{11}), and middle oils (diesel, furnace). This system is not currently receiving an extensive amount of support from the United States industrial sector because of its complex nature and high cost of implementation. However, it is being used commercially in foreign countries.

2.3.1.3 Separation--

The types of separation processes needed are dependent upon the type of coal liquefaction process used. For hydrogenation liquefaction the following separation processes may be needed:

- Flashing and condensation
- Filtration
- Centrifugation
- Solvent de-ashing
- Vacuum distillation, and
- Coking.

The separation processes associated with pyrolysis/hydrocarbonization and extraction processes include quenching the hot raw gases, separating the quenched oil from the quench liquor, and filtering the product oil.

2.3.1.4 Purification/Upgrading--

The processes used to purify and upgrade the coal-derived liquids include fractionation (distillation) and hydrotreating (hydroprocessing). Upgrading the raw coal-derived liquids by hydrotreating usually involves the addition of hydrogen in the presence of a catalyst. This process also reduces the sulfur and nitrogen content of the coal-derived fuels.

2.3.2 Status of Development

A number of processes are currently being developed to produce liquid fuels from coal. The basic idea when converting coal to liquid fuels is to obtain a product that has a higher H/C ratio than the original coal. This is achieved by either the addition of hydrogen or the removal of carbon. The H/C weight ratio of coal is about 1:(15-20) while for oil it is about 1:(6-10).

The current status of coal liquefaction processes is presented in Table 2.3-1. Operating characteristics of these processes are given in Table 2.3-2. The Fischer-Tropsch process is the only coal liquefaction process currently in commercial operation. However, four hydrogenation liquefaction processes are in advanced stages of development, and could be commercially available in the late 1980's (Ref. 2-57). These processes are: SRC-I, SRC-II, H-Coal, and EDS. Block diagrams for these four processes are presented in Figure 2.3-2. More detailed descriptions of these coal liquefaction processes can be found in the reports: Environmental Assessment Data Base for Coal Liquefaction Technology: Volume I - Systems for Fourteen Liquefaction Processes; Volume II - Detailed Discussion of H-Coal, Synthoil and Exxon Donor Solvent Processes (Ref. 2-58). A general comparison and a summary of the technical status of these four processes are shown in Table 2.3-3.

2.3.3 Applicability to and Limitations for Industrial Boilers

The applicability of coal-derived liquid fuels to industrial boiler usage is dependent upon the type of fuel desired for implementation. Because a number of different fuels are potentially available, ranging from solid to high quality liquid, a choice must be made to determine which of the many processes is most suitable for producing a fuel of the desired quality. The following section compares the applicability of different types of coal

TABLE 2.3-1. STATE OF THE ART OF COAL LIQUEFACTION PROCESSES

System Type Process Name	Developer Sponsor(s)	Present Status - PDU/* Pilot/Commercial Plant	Future Plans
Hydrogenation Systems			
Catalytic			
Bergius	I. G. Farben	• 18 commercial plants were operating in 1944 with a total capacity of 3.64 million metric tons of oil per year, Germany. • 45-TPD demonstration plant was operated in late 1950, Louisiana, MO.	No plants currently operating. Developing catalytic hydrogenation processes (H-Coal, Synthoil, Disposable Catalyst) are modern versions of Bergius process.
H-Coal	Hydrocarbon Research, Inc. Department of Energy (DOE) Ashland Oil Co. Conoco Coal Development Company Mobil Oil Corp. Standard Oil Co. (Indiana) Electric Power Research Institute (EPRI) State of Kentucky	• 2.7-TPD PDU operating, Trenton, NJ. • 546-TPD pilot plant under construction, Catlettsburg, KY.	Basic process sections of the pilot plant are due to be started up in early part of 1979. A two-year operating period is planned.
Synthoil	Pittsburgh Energy Research Center (PERC) - Division of DOE	• 2.3 and 23 kg/hr bench-scale units operating, Bruceton, PA. • 9-TPD PDU under construction, Bruceton, PA.	PDU not operating. Further plans unknown.
Disposable Catalyst	PERC-DOE	• 4.6 kg/hr bench-scale unit operating, Bruceton, PA. • 0.45-TPD PDU nearing completion, Bruceton, PA.	Runs of over 2,000 hours planned in PDU. Bench-scale unit will continue to provide additional data.
Zinc Halide	Continental Oil Co. Shell Oil Company PERC-DOE	• 0.6 kg/hr bench-scale unit operating, Library, PA. • 1.1-TPD PDU under construction, Library, PA.	Operation of PDU will begin 1978 and will continued into FY 80.
Catalytic Coal Liquefaction (Gulf CCL)	Gulf Research and Development Co. National Science Foundation (NSF)	• 2.3 kg/hr bench-scale unit operating. • 1-TPD PDU began operation in 1975, Harmarville, PA.	Conceptual design for demonstration plant being prepared.
Noncatalytic			
Solvent Refined Coal (SRC)	Pittsburgh and Midway Coal Mining Company (PAMCO) DOE	• 45-TPD pilot plant operating, Ft. Lewis, WA.	Operation of Ft. Lewis plant on SRC-I mode completed. Modified SRC-II operations continue into FY 81. A demonstration SRC-II plant processing 5454-TPD coal is planned for construction in West Virginia.
	Southern Services, Inc. EPRI DOE	• 5.5-TPD pilot plant, Wilsonville, AL.	Operation in Wilsonville plant may be extended for another three years. A demonstration SRC-I plant processing 5454-TPD coal is planned for construction in Kentucky.

(Continued)

TABLE 2.3-1. Continued

System Type Process Name	Developer Sponsor(s)	Present Status - PDU/* Pilot/Commercial Plant	Future Plans
Solvent Refined Lignite (SRL)	University of North Dakota Grand Forks Energy Research Center (GFERC), Division of DOE	• 0.55-TPD PDU operating, Grand Forks, ND.	Operating data obtained will be used to design a pilot plant.
COSTEAM	PERC-DOE GFERC-DOE	• 0.5 kg/hr bench-scale unit operating, Bruceton, PA. • 2.3 kg/hr continuous process unit under construction, Grand Forks, ND.	0.9-9 TPD PDU planned to be built at Grand Forks, ND.
Donor Solvent			
Pott-Broche	I. G. Farben	• 114-TPD demonstration plant operated from 1937-1944, Welheim, Germany.	No plants currently operating. Developing donor-solvent based extraction processes (EDS, CSF, ADL) are modern versions of Pott-Broche process.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company DOE Atlantic Richfield Co. Phillips Petroleum Co. EPRI Consortium of Japanese Companies	• 0.9 and 2.4 kg/hr bench-scale units operating, Baytown, TX. • 0.9-TPD pilot plant operating, Baytown, TX.	227-TPD pilot plant under construction near Baytown, TX. Operation will begin in 1980.
Clean Fuels From Coal (CFFC)	The Lummus Company DOE	• 13.8 kg/hr modified PDU unit operating.	Operation of PDU planned through 1978. Data obtained will be used to design and construct a pilot plant.
Consol Synthetic Fuels (CFS)	Consolidation Coal Company DOE American Electric Power Corp. Allegheny Power Corp. Flour Enggs & Consts., Inc.	• 18-TPD pilot plant operated from 1968 to 1970. Revamped since 1974 and supplemented with new process units, Cresap, WV.	Integrated operation of the plant may begin in the second half of 1978. Various methods of solids-liquid separation will be tested.
Arthur D. Little (ADL)	Arthur D. Little, Inc. Foster Wheeler Energy Corp. PERC-DOE	• Bench-scale experimental work done by ADL, Cambridge, MA. • Delayed coker PDU operating.	Further experimental studies to be continued for obtaining operating data to design PDU.
Liqui-coal	Arthur G. McKee Co.	• Bench-scale experimental work being conducted, Cleveland, OH.	Unknown.
Other Hydrogenation Processes			
Direct Hydrogenation	Rockwell International Corp. (Rocketdyne Division) DOE	• Bench-scale reactor testing was initiated in 1976. • 0.9-ton/hr reactor testing started in early 1977.	9-tons/hr reactor operated for obtaining more design data.

(Continued)

TABLE 2.3-1. Continued

System Type Process Name	Developer Sponsor(s)	Present Status - PDU/* Pilot/Commercial Plant	Future Plans
Carbonization Systems			
Char Oil Energy Development (COED)	FMC Corporation DOE EPRI	• 33-TPD pilot plant operated from 1970 to 1974, Princeton, NJ.	Pilot plant disassembled after processing 16,362 tons of coal. In a COED relevant development, COGAS process is being developed for using char from COED process to produce synthesis gas.
Clean Coke	United States Steel Engineers and Consultants, Inc. DOE	• 0.45-TPD PDU operating, Pittsburgh, PA.	Different coals will be tested in PDU through 1978. 91-TPD pilot plant design being prepared.
Coalcon	Union Carbide Corp. Chemical Construction Corp. (later withdrawn from the agreement) DOE	• 272-TPD semi-works hydrogenation pilot plant operated from 1952-1958. • 4.6 kg/hr and 18-TPD hydrocarbonization reactor scale-ups were operated in early 1960. • 2363-TPD demonstration plant was planned in 1975, New Athens, IL.	Demonstration plant project was suspended by DOE in Dec. 1976. The economics were thought to be marginal and technical problems with the fluid-bed carbonizer were thought to be bigger than first believed.
TOSCOAL	The Oil Shale Corp. (TOSCO)	• 23-TPD pilot plant originally built to process shale was tested with coal, Golden, CO.	No future plans known.
Flash Pyrolysis	Occidental Research Corp. (ORC) DOE State of Kentucky	• 3.2 TPD PDU operated for testing the process. This PDU has currently been revised, La Verne, CA. • 2.8-TPD second PDU started operating in 1977, La Verne, CA.	The PDU operations will continue through 1978. Tar hydrotreating will also be tested in PDU. A detailed design for a 227-TPD pilot plant is being prepared.
Extraction			
Supercritical Gas Extraction (SGE)	Coal Research Establishment (CRE) British National Coal Board (NCB)	• Experimental work done over a number of years, Stoke Orchard, U.K. • 20 kg/hr PDU commissioned by CRE.	Air Products and Chemicals, Inc. has been contracted to examine the engineering of the process and to investigate its commercial potential.
Catalytic Synthesis			
Fischer-Tropsch	Lurgi-Ruhrchemie (ARGE), Oberhausen, Germany	• 7272-TPD commercial plant operated by South African Coal, Oil and Gas Corp. (SASOL) of the Republic of South Africa, Sasolburg, South Africa.	SASOL-II plant under construction in South Africa at the new town Secunda, located in the Transvaal province.
	PERC-DOE	• Frame-sprayed catalyst system is being developed.	Data obtained will be used to design a pilot plant.

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* All tons are in metric tons.

TABLE 2.3-2. OPERATING CHARACTERISTICS OF COAL LIQUEFACTION PROCESS

System Type Process Name	Characteristics of Coal Pretreatment Operation			Characteristics of Liquefaction Reaction Operation		
	Acceptable Coal Type	Cool Feeding System	Acceptable Coal Size, mm (moisture content, %)	Type of Reaction Vessel	Operating Pressure MPa	Operating Temperature °C
Hydrogenation Systems						
Bergius	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	0.25 (1.5)		68.94	482
H-Coal	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	<0.35 (0)	Ebullated bed (catalytic)	20.20	454
Synthoil	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	60%<0.149 (<3.0)	Fixed bed (catalytic)	13.73-27.46	449
Disposable Catalyst	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	60%<0.149	Stirred tank	11.5-23.0	449
Zinc Halide	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	--	Molten bed (catalytic)	10.49-20.98	400
Gulf CCL	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	--	Fixed bed (catalytic)	13.72	482
SRC	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	<3.51 (<3.0)	Tubular flow	10.29	454
SRL	Lignite	Coal/solvent slurry	--	Tubular flow	17.35	510
COSTEAM	Lignite	Coal/solvent slurry	<0.149 (20 to 30)	Stirred bed	27.44	427

(Continued)

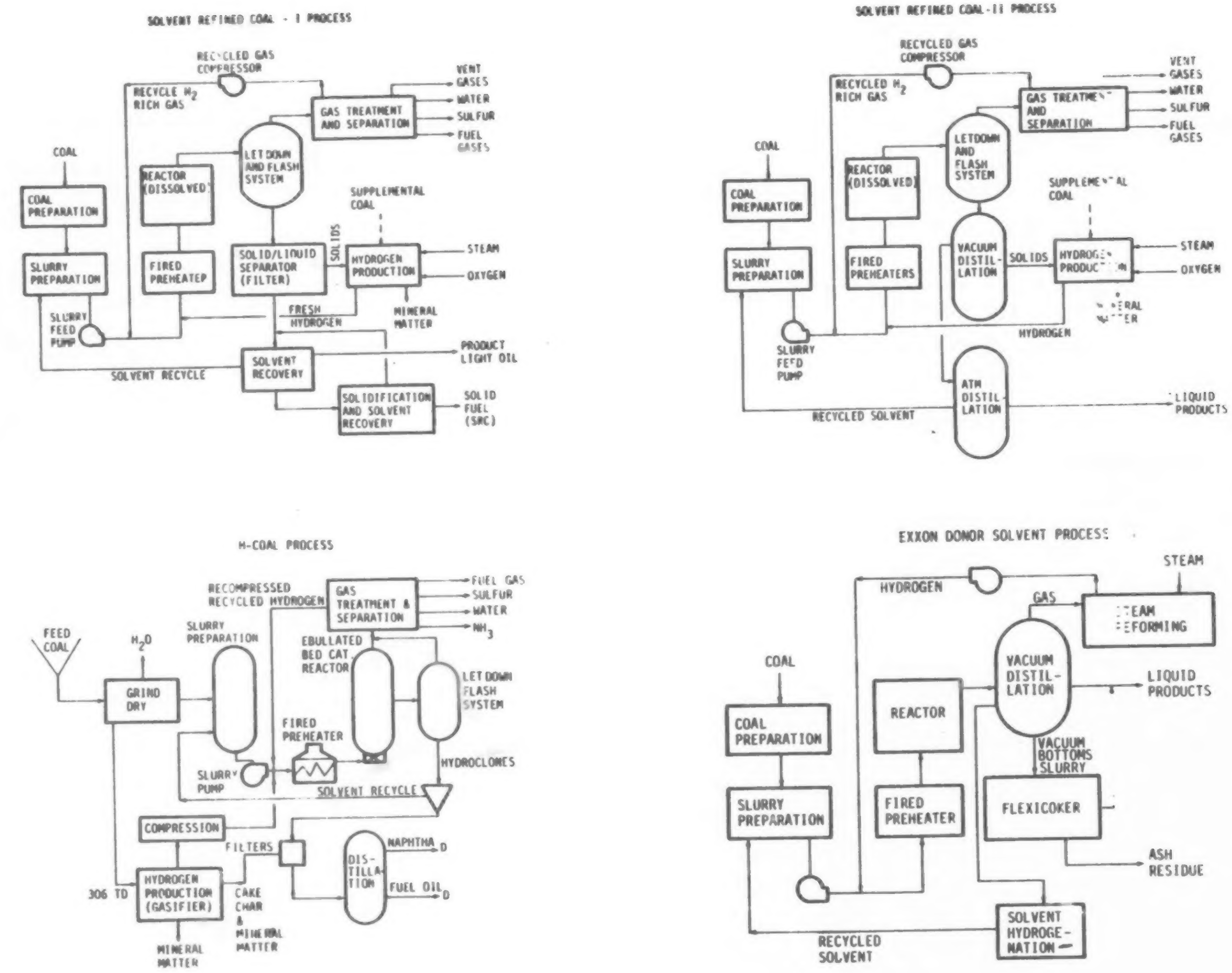
TABLE 2.3-2. Continued

System Type	Characteristics of Coal Pretreatment Operation			Characteristics of Liquefaction Reaction Operation		
Process Name	Acceptable Coal Type	Coal Feeding System	Acceptable Coal Size, mm (moisture content, %)	Type of Reaction Vessel	Operating Pressure MPa	Operating Temperature °C
Pott-Broche	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	95%<0.084 (0.5)	Tubular flow	10.10	430
EDS	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	<0.54 (--)	Tubular flow	13.72	449
CFFC	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	--	Multistage ebullated bed (catalytic)	14.60	427
CSF	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	<1.19 (1.0)	Stirred tank	0.98-2.74	407
ADL	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	80%<0.074 (--)	Stirred tank	0.79	399
Liqui-coal	Bituminous Sub-bituminous Lignite	Coal/solvent slurry	<0.149 (2.0)	Tubular flow	6.86	427
Direct Hydrogenation	Bituminous Sub-bituminous Lignite	Pneumatic	<0.074 (--)	Entrained bed	6.35	982
Extraction Systems						
SGE	Bituminous Sub-bituminous Lignite	Lock hoppers	<0.074 (--)	--	10.1	399

(Continued)

TABLE 2.3-2. Continued

System Type Process Name	Characteristics of Coal Pretreatment Operation			Characteristics of Liquefaction Reaction Operation		
	Acceptable Coal Type	Coal Feeding System	Acceptable Coal Size, mm (moisture content, %)	Type of Reaction Vessel	Operating Pressure MPa	Operating Temperature °C
Carbonization Systems						
COED	Bituminous Sub-bituminous Lignite	Pneumatic	3.51 (6.0)	Multistage fluidized bed	0.12	316-816 (4 stages)
Clean-coke	Sub-bituminous (high volatile)	Coal/solvent slurry	0.149-3.51 (--)	Fluidized bed	0.57-0.82 (carbonizer)	732
				--	16.82-25.44 (hydrogenation)	469
Coalcon	Bituminous Sub-bituminous Lignite	Lock hoppers	50%, 0.149 (1.0)	Fluidized bed	3.28	560
TOSCOAL	Sub-bituminous	Pneumatic	12.7	Fluidized bed	0.08	482
Flash Pyrolysis	Bituminous Sub-bituminous Lignite	Pneumatic	0.074 (--)	Entrained bed	0.08	593
Catalytic Synthesis						
Fischer-Tropsch (gasification)	Depends upon the gasifier used	Depends upon the gasifier used	Depends upon the gasifier used	Moving bed	2.30	593-927



SOLVENT REFINED COAL - I PROCESS

COAL
COAL PREPARATION
SLURRY PREPARATION
SLURRY FEED PUMP
RECYCLE H2 RICH GAS
RECYCLED GAS COMPRESSOR
REACTOR (DISSOLVED)
FIRED PREHEATER
LET DOWN AND FLASH SYSTEM
GAS TREATMENT AND SEPARATION
VENT GASES
WATER
SULFUR
FUEL GASES
SUPPLEMENTAL COAL
SOLID/LIQUID SEPARATOR (FILTER)
SOLIDS
HYDROGEN PRODUCTION
STEAM
OXYGEN
MINERAL MATTER
FRESH HYDROGEN
SOLVENT RECOVERY
SOLVENT RECYCLE
PRODUCT LIGHT OIL
SOLIDIFICATION AND SOLVENT RECOVERY
SOLID FUEL (SRC)

SOLVENT REFINED COAL-II PROCESS

COAL
COAL PREPARATION
SLURRY PREPARATION
SLURRY FEED PUMP
RECYCLED H2 RICH GAS
RECYCLED GAS COMPRESSOR
REACTOR (DISSOLVED)
FIRED PREHEATERS
LETDOWN AND FLASH SYSTEM
GAS TREATMENT AND SEPARATION
VENT GASES
WATER
SULFUR
FUEL GASES
SUPPLEMENTAL COAL
VACUUM DISTIL-LATION
SOLIDS
HYDROGEN PRODUCTION
STEAM
OXYGEN
MINERAL MATTER
HYDROGEN
ATM DISTIL-LATION
RECYCLED SOLVENT
LIQUID PRODUCTS

H-COAL PROCESS

FEED COAL
H2O
GRIND DRY
SLURRY PREPARATION
SLURRY PUMP
RECOMPRESSED RECYCLED HYDROGEN
GAS TREATMENT & SEPARATION
FUEL GAS
SULFUR
WATER
NH3
EBULLATED BED CAT REACTOR
FIRED PREHEATER
LET DOWN FLASH SYSTEM
HYDROCLONES
SOLVENT RECYCLE
COMPRESSION
FILTERS
DIS-TILLA-TION
NAPHTHA
FUEL OIL
306 TD
HYDROGEN PRODUCTION (GASIFIER)
CAKE CHAR & MINERAL MATTER
MINERAL MATTER

EXXON DONOR SOLVENT PROCESS

COAL
COAL PREPARATION
SLURRY PREPARATION
HYDROGEN
REACTOR
FIRED PREHEATER
GAS
STEAM
STEAM REFORMING
VACUUM DISTIL-LATION
LIQUID PRODUCTS
VACUUM BOTTOMS SLURRY
FLEXICOKER
ASH RESIDUE
SOLVENT HYDROGE-NATION
RECYCLED SOLVENT

Figure 2.3-2. Flow diagrams of advanced liquefaction processes.

TABLE 2.3-3. GENERAL COMPARISON AND RELATIVE TECHNICAL STATUS OF FOUR HYDROGENATION LIQUEFACTION PROCESSES

Technical Status	SRC-I	SRC-II	H-Coal*	EDS*
PDU Scale of Operations (metric tons/day, coal)	0.9	0.9	0.27	0.27
Size of Pilot Plant (metric tons/day, coal)	a) 45 b) 5.5	27.3	227.3-545.4	227.3
Pilot Plant Scheduled Operation	a) Late 1974 b) Mid-1976	Mid-1977	April 1979	January 1980
Fuel Types	Refined coal (solid fuel)	Distillate oil	Distillate oil residual oil	Distillate oil
Coal Feed	Eastern Western	Eastern (high pyrites only)	Eastern Western	Eastern Western
General Comparisons				
Fuel Flexibility	Developed to produce substitute solid boiler fuel only	Poor	Superior	Good
Reactor Operating Severity	Moderate	High	Moderate	Low
Process Scale-up Risk	Moderate	High	Moderate*	Low*
Number of New Components and Design of Commercial Equipment	High	High	Low*	Low*
Reactor Complexity	Moderate	Moderate	Moderate	Low
Fuel Utilization, Combustion				
Raw Product, Stability, Compatibility	Moderate	Poor	Poor	Poor
Combustion Experience	Moderate	Limited	Limited	Limited

*Assumes successful operation of pilot plant in 1979-1980.

liquefaction fuels to several boiler configurations in use today. The fuel characteristics of various coal-liquids are also discussed.

Commercial coal liquefaction facilities will produce fuels in much larger quantities than are required by any one industrial boiler. Therefore the liquefaction plant will be considered as an off-site supplier of fuel.

2.3.3.1 Use of Coal-Derived Solid or Liquid Fuels in Boilers--

The products from the four coal liquefaction processes of concern are a solid fuel and a liquid fuel or chemical feedstock. The solid fuel produced by the SRC-I process cannot be used in all industrial coal-fired boilers. However, the liquid fuels can be used in oil-fired boilers.

The solid fuel from the SRC-I process cannot be used in conventional stoker boilers, which comprise about 11 percent of industrial and commercial boilers currently in use (Ref. 2-59). The spreader stoker, for example, burns large size coal particles both above and on the grate. Since the solids from SRC-I have a low melting point (approximately 155°C), they will melt on the grate before they are combusted. The melted SRC-I will then pass through the grate into the plenum chamber and be removed with the ash. This results in very inefficient combustion.

Four percent of industrial and commercial boilers are pulverized coal boilers (Ref. 2-59). Some of the pulverized coal boilers may be retrofitted to burn SRC-I fuel. However, preliminary results from SRC-I handling and burning tests indicate that some modifications may be required for the operation of the fuel handling and storage equipment, pulverizers, boiler burners and the combustion process itself. Such modifications for the ball-and-race mill might include reduction of the loading force and/or race speed and change in the operating procedure to minimize holdup of SRC in the pulverizer (Ref. 2-60). For combustion, the dual-register burner may have to be redesigned with a water-cooled coal nozzle (Ref. 2-61).

Approximately 34 percent of industrial and commercial boilers in operation are oil-fired (Ref. 2-59). It is anticipated that coal liquids may replace a portion of the petroleum-based fuel oil-fired industrial boilers with modifications in the combustion process. Only minor modifications may be required in the handling and storage of coal liquids.

Coal-derived heavy liquid fuels can be burned in boilers using petroleum-based residual oils (Nos. 4, 5, and 6 fuel oils). These boilers presently comprise 20 percent of industrial and commercial boilers in operation (Ref. 2-59). Analyses (Ref. 2-60) of petroleum-based fuel oils (Nos. 1, 2, 4, 5, and 6) are given in Table 2.3-4. Fuels from both the H-Coal (fuel-oil mode) and SRC-II processes may be used to replace residual fuel oil. Utilization of EDS liquid fuel as residual fuel will depend on the extent of hydrotreatment.

Fourteen percent of industrial and commercial boilers use No. 2 fuel oil (Ref. 2-59). The EDS process produces a distillate fuel which can be substituted for petroleum-based No. 2 fuel oil. Mild hydroprocessing may be required for H-Coal and SRC-II liquids to produce a product similar to No. 2 fuel oil.

2.3.3.2 Coal-Liquids Fuel Characteristics--

Coal liquids are analogous to petroleum residual fuels. However, analysis of these coal liquids shows that their chemical compositions and certain physical properties are sufficiently different to require possible handling and combustion modifications before use in industrial boilers. Coal liquids are more aromatic and their carbon/hydrogen ratios are considerably higher than those of petroleum crudes and fuels. Moreover, the concentrations of heteroatoms such as nitrogen and oxygen are much higher. Coal liquids containing residual fractions high in asphaltenes are incompatible when blended with typical petroleum fuels. The sulfur contents of both coal liquids and petroleum residuals can vary considerably. The physical properties of coal liquids are also quite variable; they depend on the type of coal used and the

TABLE 2.3-4. RANGE OF ANALYSES OF FUEL OILS

Grade of Fuel Oil	No. 1	No. 2	No. 4	No. 5	No. 6
Weight, Percent					
Sulfur	0.01-0.5	0.05-1.0	0.2-2.0	0.5-3.0	0.7-3.5
Hydrogen	13.3-14.1	11.8-13.9	(10.6-13.0)*	(10.5-12.0)*	(9.5-12.0)*
Carbon	85.9-86.7	86.1-88.2	(86.5-80.2)*	(86.5-89.2)*	(86.5-90.2)*
Nitrogen	Nil-0.1	Nil-0.1			
Ash			0-0.1	0-0.1	0.01-0.5
Gravity					
Deg API	40-44	28-40	15-30	14-22	7-22
Specific	0.825-0.806	0.887-0.825	0.966-0.876	0.972-0.922	1.022-0.922
kg/m^3	110.06-107.49	118.39-110.06	128.80-116.95	129.76-123.03	136.33-123.03
Pour Point, °C	0 to -45.6	0 to -40	-23.3 to 10	-23.3 to 26.7	-9.4 to 29.4
Viscosity					
Centistokes @ 37.8°C	1.4-2.2	1.9-3.0	10.5-65	65-200	260-750
SUS @ 37.8°C		32-38	60-300		
SSF @ 50°C				20-40	45-300
Water & Sediment, Vol. %		0-0.1	tr to 1.0	0.05-1.0	0.05-2.0
Heating Value (MJ/kg)	45.75-46.19	44.59-45.94	42.52-45.12	42.1-44.2	40.5-44.17
(calculated Btu/lb) gross	(19,670-19,860)	(19,170-19,750)	(18,280-19,400)	(18,100-19,020)	(17,410-18,990)

*Estimated.

liquefaction process employed. Petroleum residuals have higher concentrations of nickel and vanadium (10 to 500 ppm) than coal liquids (Ref. 2-61). However, coal-derived liquids contain more iron (∿10 to 60 ppm).

A comparison of physical and chemical characteristics of a typical petroleum No. 2 fuel oil and coal liquids is given in Table 2.3-5. Coal liquids (Ref. 2-62) in the table are recycled SRC solvent (obtained from SRC-I and considered here to be analogous to SRC-II liquid fuel), H-Coal distillate (obtained by operating H-Coal process in fuel oil mode without hydroprocessing), and EDS distillate. In comparison, No. 2 fuel oil exhibits higher hydrogen content than coal liquids. Also, nitrogen content in these coal-liquid fuels is considerably higher than No. 2 fuel oil. According to the fuel characteristics shown in one source (Ref. 2-64), the ash content of SRC-II liquid oil was between No. 2 and No. 6 fuel oil. Viscosity of coal liquid fuels is high and the coal liquids must be heated for proper atomization in the boiler.

The properties of SRC-I (solid fuel) are compared with those of base coals in Table 2.3-6. The sulfur content of the base coals is reduced to 0.8 and 0.1 percent respectively for the high sulfur Kentucky and low sulfur Wyoming coals. The ash content in both cases was reduced to 0.3 percent or less. The hydrogen content, as expected, is somewhat higher for the SRC-I as compared with the base coal. Some nitrogen enrichment has occurred, giving SRC-I a slightly higher nitrogen content than the base coal.

Oxygen contents of the SRC-I are about half the value of the base coal. The Hardgrove grindability index for SRC-I is 145 (Ref. 2-66), indicating that the power required to pulverize should be significantly less than that required for base coal, which has a Hardgrove grindability index of about 50 to 100. As mentioned earlier, some modifications will be required to pulverize SRC-I. Babcock and Wilcox Company's E-21 ball-and-race mill (Ref. 2-67) and C-E Deep Bowl Mill No. 271 (Ref. 2-68) have shown satisfactory performance in pulverizing SRC-I.

TABLE 2.3-5. PROPERTIES OF NO. 2 FUEL OIL AND COAL LIQUIDS

Fuel	Typical Petroleum No. 2 Fuel Oil Ref. 2-62	Raw SRC-I Recycled Solvent* Ref. 2-62	Raw H-Coal Distillate Ref. 2-62	Raw EDS 205-425°C Distillate Ref. 2-63
Elemental Composition, wt %				
Hydrogen	12.9	7.4	9.1	10.0
Sulfur	0.093	0.37	0.10	0.3
Nitrogen	0.008	0.62	0.39	0.2
Oxygen	<0.01	3.9	1.5	1.9
Trace Contaminants, ppm wt				
Titanium	<1.0	20.0		
Sodium	0.55	0.39	0.59	
Potassium	0.37	0.19	0.08	
Calcium	0.17	0.35	0.14	
Vanadium	<0.1	0.9	0.1	
Lead	-	0.9	<1.0	
Iron	0.2	61.0	10.3	
Chloride	-	35.0		
Physical Properties				
Gravity, °API	33.6	5.3	14.7	18.0
Aromatic Carbon, %	19.0	74.0	55.0	-
Flash Point, °C	67.2	82.2	90.5	91.1
Heat of Combustion MJ/kg (Btu/lb)	45.4 (19500)	39.4 (16920)	42.1 (18080)	44.2 (19000)
KV, CS at (43.3°C)	2.61	5.79	3.0	SUS @ 37.8°C = 38
KV, CS at (98.9°C)	1.09	1.48		

*SRC-I recycled solvent is considered analogous to SRC-II liquid fuel.

TABLE 2.3-6. PROPERTIES OF SRC-I AND BASE COALS

Ultimate Analysis wt %	Kentucky High Sulfur Coal	Kentucky SRC-I	Wyoming Low-Sulfur Coal	Wyoming SRC-I
Moisture	2.2	1.1	6.5	0.2
Ash	10.5	0.3	7.3	0.2
C	70.1	86.1	63.4	88.4
H	4.9	5.7	4.5	5.5
N	1.4	1.7	1.0	1.3
S	3.5	0.8	0.6	0.1
O	7.3	4.3	16.7	4.3
Heating Value, MJ/kg	29.4	36.2	23.9	36.6
(Btu/lb)	(12,600)	(15,600)	(10,300)	(15,700)

Source: Ref. 2-65

2.3.4 Factors Affecting Performance

In coal liquefaction systems, sulfur and nitrogen present in coal react with hydrogen to form hydrogen sulfide (H_2S) and ammonia (NH_3), respectively. The system operating characteristics play an important role in the efficiency of these conversions. The primary variables that could affect the conversions are:

- Type of system
- Coal characteristics
- Operation variables
  - Hydrogen consumption
  - Reactor space velocity
  - Reaction temperature and pressure
- Hydroprocessing of the raw product

2.3.4.1 Type of System--

In the coal liquefaction reactor, coal is solubilized in the solvent and is further hydrogenated by the addition and/or transfer of hydrogen to remove sulfur, nitrogen, and oxygen. Liquid and gaseous products are produced by hydrocracking. The severity of the reaction is measured by the extent to which hydrogenation and hydrocracking reactions proceed.

In the SRC-I solid system, the reaction severity is low. The hydrogenation and hydrocracking reactions result in the desired product having respective sulfur and ash contents of about 0.7 and 0.1 percent. Nitrogen contents may be slightly increased.

In the SRC-II (liquid) system, a portion of the reactor effluent slurry is recycled back to the reactor. The mineral residue in the product slurry is believed to act as a catalyst and enhance hydrocracking reactions. More hydrogen is transferred, and the main product is a distillate fuel with sulfur and nitrogen contents of about 0.3 and 0.9 percent, respectively. Lower sulfur and nitrogen contents in the final product signify larger amounts of hydrogen sulfide and ammonia generated in the reactor, and a reaction severity greater than in the SRC-I system.

In the H-Coal reactor, hydrogenation and hydrocracking reactions are activated by the addition of a catalyst, which helps minimize hydrogen consumption. The reactions are sufficiently complete to allow production of fuel oil distillate (204^{+}°C oil) with a sulfur content of 0.6 percent. Operating with a higher consumption of hydrogen in the reactor, the system produces syncrude with a sulfur content of 0.2 percent. The reaction severity is higher than in the SRC system.

In the EDS liquefaction reactor, the solubilized coal is hydrocracked to liquid products, which are separately hydrogenated in a catalytic reactor to reduce the sulfur and oxygen content. The process produces a variety of liquid products with sulfur content of less than 0.1 percent. The reaction severity in the liquefaction reactor is low, but the combined severity of the liquefaction reactor and hydrogenation reactor is quite high.

In summary, the effect of system type in producing lower sulfur and nitrogen content fuels is that higher sulfur and nitrogen removal efficiencies are obtained with higher severity of hydrogenation and hydrocracking reactions. Table 2.3-7 shows the variation in hydrogen consumption and sulfur content of the product with the type of liquefaction system.

TABLE 2.3-7. HYDROGEN CONSUMPTION AND SULFUR CONTENT OF COAL LIQUIDS IN LIQUEFACTION SYSTEMS

	SRC-I (solid) (Ref. 2-56)	SRC-II (liquid) (Ref. 2-56)	H-Coal (fuel oil) (Ref. 2-69)	EDS (Ref. 2-70)
H_2 consumption, % by wt of dry coal*	2.20	4.30	3.88	3.05
% sulfur in 205^{+}°C oil	0.70	0.30	0.58	0.10

* Coal type: Illinois No. 6, sulfur content 3.5%.

2.3.4.2 Coal Characteristics--

Coal contains organic sulfur (bound in organic compounds) and mineral sulfur (as pyrites and other iron compounds). Mineral sulfur is easily removed during the hydrogenation reaction with very little hydrogen consumption. A summary of data on low-sulfur, low-ash SRC-I product from various coals in the Wilsonville pilot plant is presented in Table 2.3-8.

Nitrogen removal is more difficult than sulfur removal because the reaction between nitrogen and hydrogen does not take place easily. Nitrogen removal from the feed coal ranges from approximately zero to 40 percent. About half of the nitrogen removal goes into the production of ammonia.

TABLE 2.3-8. SULFUR IN FEED COALS AND FLUE GASES FROM COMBUSTION OF FEED COALS AND SRC PRODUCTS

	Sulfur in Feed Coal[a] wt %			Flue Gas SO_2, $ngSO_2/J$ ($lb/10^6$ Btu)		
	Total	Organic	Mineral	Feed Coal[b]	SRC Product[e]	Sulfur Removal[c] %
Smith & Roland, Powder River, Belle Ayr	0.81	0.71	0.10	590 (1.36)	80 (0.18)	87
Emergy Field/Emery	0.94	0.72	0.22	610 (1.41)	180 (0.41)	71
Illinois No. 6/Monterey	3.89	2.53	1.36	3030 (7.04)	480 (1.12)	84
Illinois No. 6/Burning Star	3.10	2.00	1.10	2520 (5.85)	340 (0.78)	87
Pittsburgh/Loveridge	2.79	1.76	1.03	2060 (4.78)	460 (1.07)	78
Kentucky No. 9 and No. 14 (A[d])/Colonial	4.04	1.01	3.03	3130 (7.27)	220 (0.52)	93
Kentucky No. 9 and No. 14 (B[d])/Colonial	3.66	1.50	2.16	3010 (7.24)	300 (0.70)	90
Indiana V/Old Ben	3.21	1.25	1.96	2530 (5.88)	320 (0.75)	87

[a] Coal contains organic sulfur (bound in organic compounds) and mineral sulfur (as pyrites and other iron compounds). The total sulfur is the sum of organic and mineral sulfur.

[b] Run-of-mine coal.

[c] Takes no credit for low-sulfur liquids and gases produced. Calculated from flue gas SO_2 values.

[d] Wide range of coal composition from the same mine.

[e] SRC product made from designated feed coal.

Source: Ref. 2-71.

2.3.4.3 Operation Variables--

The effects of operating variables such as hydrogen consumption, space velocity, temperature, and pressure on removal of sulfur and nitrogen are considered jointly since space velocity, temperature, and pressure all affect hydrogen consumption. An increase in reaction temperature and pressure causes increased reaction rates, along with an increase in hydrogen consumption. An increase in residence time (decrease in space velocity) also increases the consumption of hydrogen. Since hydrogen consumption is influenced by all these variables, it can be used as an indicator of the sulfur and nitrogen removal.

A. Desulfurization--Increased hydrogen consumption indicates an increase in sulfur removal. The effect of hydrogen consumption on the sulfur content in SRC produced from various coals is shown in Figure 2.3-3 (Ref. 2-72). Reactor temperature and residence time have greater effects on sulfur removal than reactor pressure (Ref. 2-73). At low temperatures, the relationship between sulfur removal and hydrogen consumption is approximately linear. However, as the temperature is increased, more sulfur is removed but at a lower rate. Sulfur removal can also be increased by increasing reactor pressure and residence time.

B. Denitrogenation--During coal hydrogenation, nitrogen reacts with hydrogen to form ammonia. An increase in temperature increases hydrogen consumption; however, nitrogen removal at higher temperatures does not significantly change. Figure 2.3-4 summarizes calculated nitrogen contents against the operating conditions employed (Ref. 2-72). The lines drawn for the effect of the variables represent averaged results. It appears that the actual effect of these variables is small, on the order of 0.1 wt percent nitrogen content change for change in pressure from 6.86 to 10.49 MPa, or a change in coal rate from 400 to 799 kilograms per hour per cubic meter of contactor.

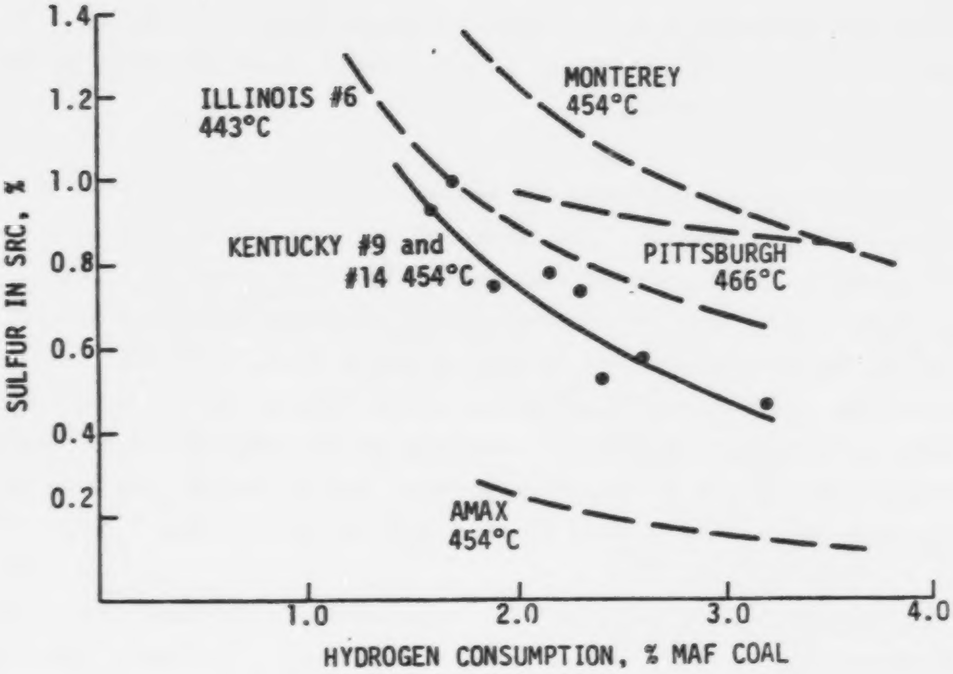


Figure 2.3-3. Effect of hydrogen consumption upon sulfur in filtered SRC (Ref. 2-72).

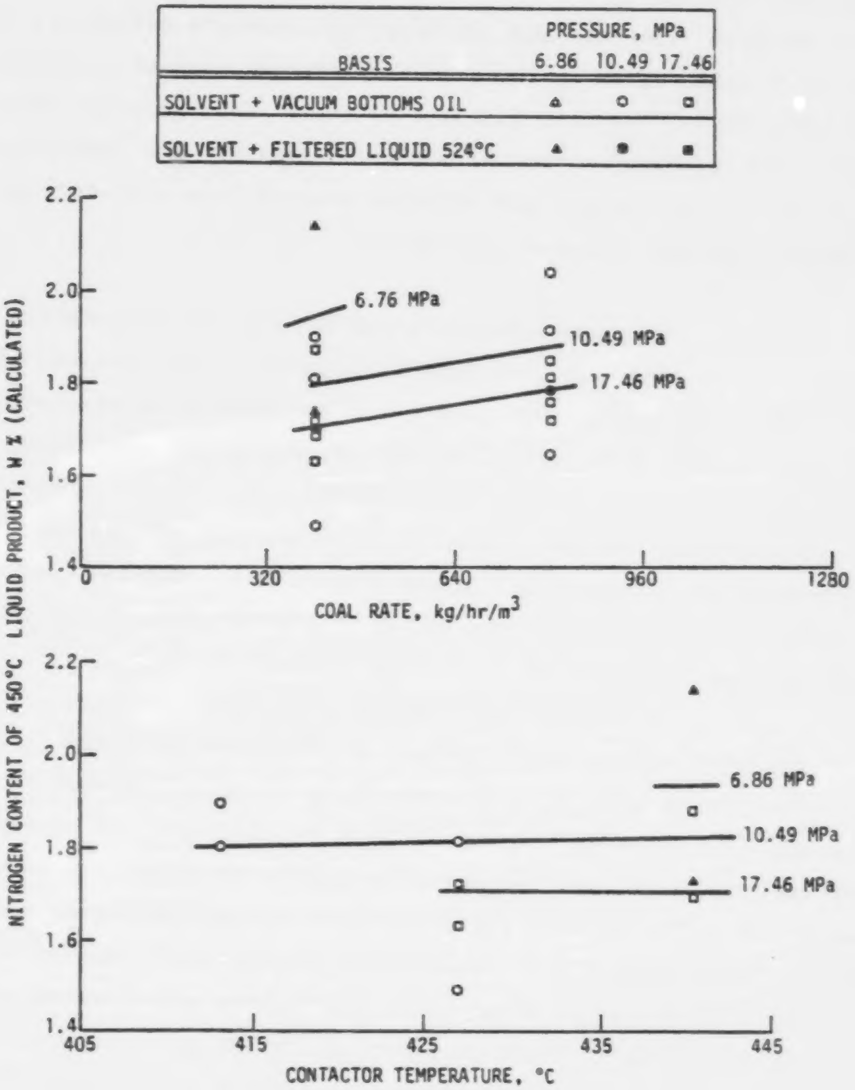


Figure 2.3-4. Variation of nitrogen content of SRC-I liquids with changes in temperature and coal-feed rate (Ref. 2-72).

2.3.4.4 Hydroprocessing--

Raw coal-derived liquid fuels differ from petroleum-derived fuels in that they are very aromatic and, as such, are hydrogen deficient. Hydroprocessing of these fuels to increase the hydrogen content consists of catalytically adding chemical hydrogen to fuel. Hydrogenation also decreases the sulfur and nitrogen content of the coal liquids. The hydroprocessing variables that affect sulfur and nitrogen removal from coal-derived liquids are catalyst type and hydrogen consumption.

A. Catalyst Type--Mobil Research and Development Corporation (Ref. 2-73) has evaluated several catalysts for upgrading coal-derived liquids to turbine fuels. The evaluations were based on bench-scale apparatus using coal-derived liquids from SRC and H-Coal pilot plants. Commercial CoMo/Al_2O_3 catalyst (Cyanimid HDS-1441A) and NiMo/Al_2O_3 catalyst (Ketjen 153S) have shown good performances. Solvent hydrogenation studies performed by Exxon Research and Engineering Company (Ref. 2-74) indicated that the denitrogenation capability of the nickel-molybdate catalyst appears to be somewhat greater than that of the cobalt-molybdate catalyst.

B. Hydrogen Consumption--Hydrogen consumption for a hydroprocessing reaction depends on the degree of heteroatom (O, N, S) removal desired. Lower space velocities and higher temperatures increase the hydrogen consumption, which in turn increases the degree of heteroatom removal. Figures 2.3-5 and 2.3-6 show the effect that average operating temperature has on desulfurization and denitrogenation during hydrotreating operations (Ref. 2-74). It should be noted that temperatures and space velocities reported by Exxon Research and Engineering Company are relative values; that is, they are reported as functions of reference values (R.V.) defined by Exxon. The feedstock for the experimental runs, characterized as "simulated heavy solvent" (Ref. 2-74), consists of multipass spent solvent from the EDS pilot plant. Table 2.3-9 represents mild, moderate, and severe hydroprocessing conditions for recycled SRC solvent and H-Coal

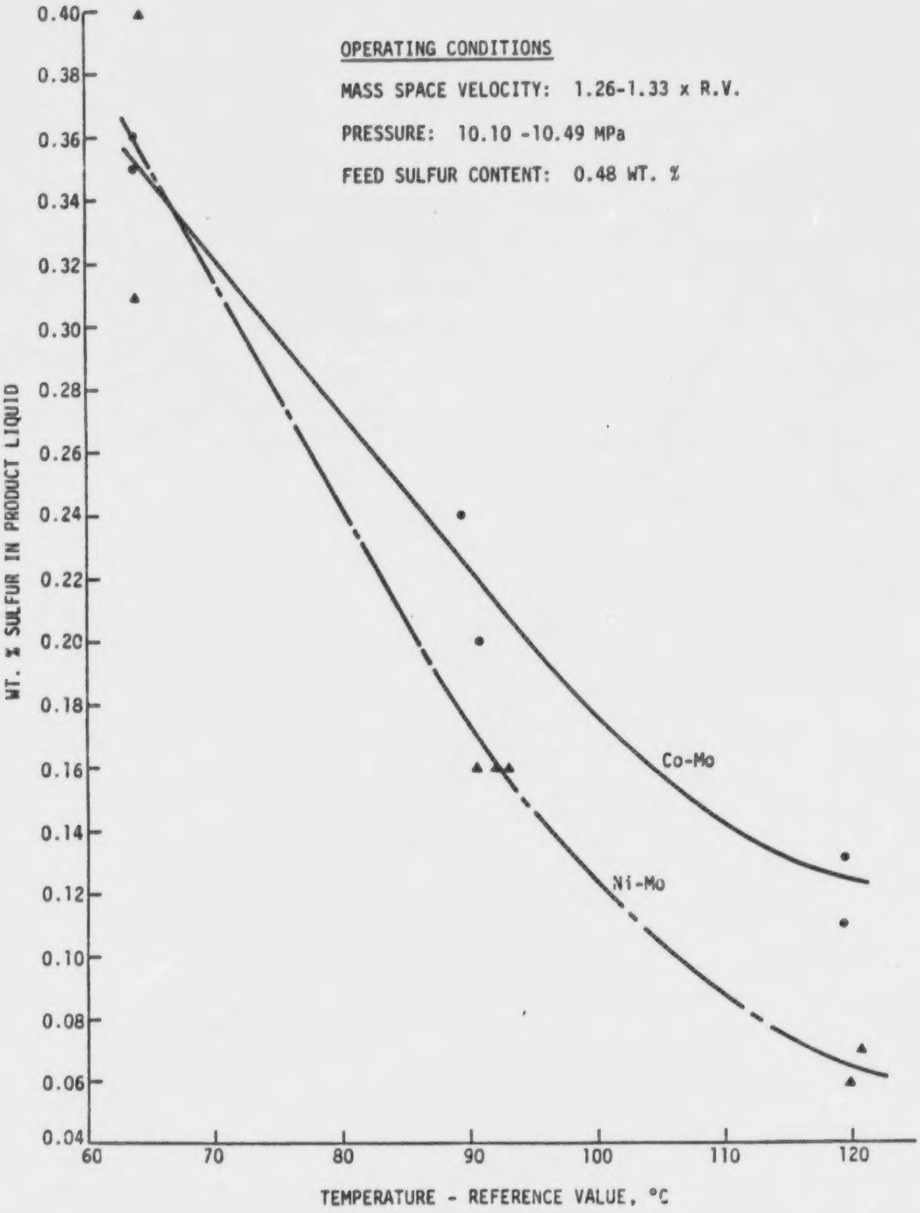


Figure 2.3-5. Effect of temperature on sulfur removal (Ref 2-74).

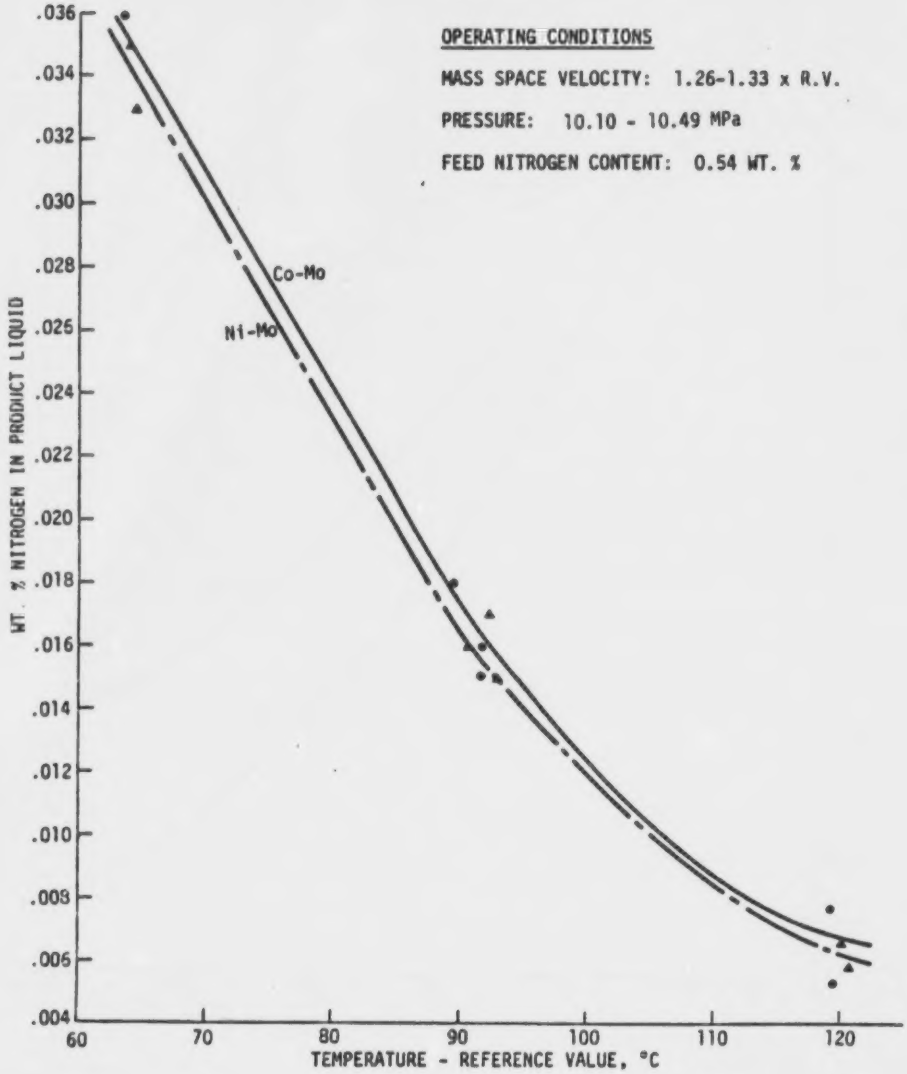


Figure 2.3-6. Effect of temperature on nitrogen removal (Ref. 2-74).

TABLE 2.3-9. HYDROPROCESSING CONDITIONS AND YIELDS FOR UPGRADING SRC RECYCLE SOLVENT AND H-COAL DISTILLATE

Coal Liquid	SRC Recycle Solvent				H-Coal Distillate		
Hydroprocessing Conditions:*	Feed	Mild	Moderate	Severe	Feed	Mild	Severe
Pressure, MPa	-	17.2	17.2	17.2	-	10.3	17.2
LHSV	-	2.9	0.8	0.4	-	2.9	0.5
Temperature, °C	-	359	377	381	-	369	372
H_2 Consumption (m^3/m^3 oil)	-	202.9	402.3	498.4	-	101.5	307.9
(SCF/Bbl)		(1140)	(2260)	(2800)		(570)	(1730)
Yields, wt %							
C_1-C_3 Gas	-	0.16	0.11	0.42	-	0.04	0.20
C_4	-	0.22	0.07	0.22	-	0.01	0.10
C_5	-	0.11	0.05	0.12	-	0.03	0.06
C_6 + Liquid	-	98.90	98.34	98.05	-	99.60	100.21
H_2S	-	0.34	0.38	0.39	-	0.13	0.14
NH_3	-	0.22	0.62	0.73	-	0.18	0.46
H_2O	-	1.72	3.73	4.17	-	0.90	1.58
Heteroatom Removal, wt %							
Sulfur	-	85	97	98	-	95	99
Oxygen	-	39	85	95	-	54	93
Nitrogen	-	30	83	96	-	40	99
Total Liquid Product Properties							
Gravity, °API	5.3	13.0	19.5	23.4	17.1	21.3	28.0
Hydrogen, wt %	7.4	8.9	10.3	11.0	9.8	10.6	12.2
Sulfur, wt %	0.37	0.06	0.01	0.01	0.13	0.007	<0.002
Nitrogen, wt %	0.62	0.44	0.11	0.02	0.38	0.23	<0.005
Oxygen, wt %	3.9	2.4	0.6	0.2	1.5	0.7	0.1

* Catalysts: American Cyanimid HDS-1141A (CoMo) for SRC Recycle Solvent and Ketjen 153S (NiMo) for H-Coal Distillate.

Source: Ref. 2-61

distillate (Ref. 2-61). The sulfur removal for recycled SRC solvent is greater than 85 percent for all three conditions, while nitrogen and oxygen removal varied from 30 to 96 percent. For H-Coal distillate, the sulfur removal was greater than 95 percent, while oxygen and nitrogen removal varied from 30 to 99 percent.

2.3.5 System Performance

There is limited storage and combustion data on coal liquefaction products from any of the processes. The Ft. Lewis, Washington, SRC-I pilot plant with a capacity of 45 metric tons per day, produced a 2725 metric ton sample of solid SRC fuel for combustion testing at the 22 MW_e Plant Mitchell power station of the Georgia Power Company. The combustion tests were performed in the second quarter of 1977. Small-scale tests on home heating units and industrial boilers, and some limited laboratory tests have been performed with SRC-II, H-Coal, and EDS liquid fuels. Large-scale boiler tests, using SRC-II fuel from several coals at the Ft. Lewis pilot plant, are being planned. Tests on H-Coal and EDS fuel products will be initiated in 1979 and 1980, respectively (Ref. 2-75).

2.3.5.1 Emission Reductions--

The results of combustion tests performed on coal-derived solid and liquid fuels are discussed in this section. These results were primarily concerned with sulfur dioxide, nitrogen oxides, and particulate matter emissions.

A. Sulfur Dioxide--When combusted, most of the coal-derived liquid and solid fuels will meet an SO_2 emission level of 520 ng SO_2/J (1.2 lb/10^6 Btu) without any post-combustion controls (flue gas desulfurization). SRC-I fuel with a heating value of 37 MJ/kg (16,000 Btu/lb) will meet this SO_2 emission level if its sulfur content is less than 0.96 percent. For an SRC-I fuel with a heating value of 35 MJ/kg (15,000 Btu/lb) the sulfur content will need to be less than 0.90 percent. The SRC-I produced during the Ft. Lewis pilot plant runs had a typical sulfur content of 0.7 percent and when burned in the Plant Mitchell facility SO_2 emissions ranged from 400 to 460 ng/J (0.9-1.1 lb/10^6 Btu) (Ref. 2-76, 2-77). This is approximately an 80-85 percent reduction of the emissions that would arise from direct-firing of the original coal feed. SRC-I produced from other high sulfur coals, however, may produce higher SO_2 emissions and will require flue gas desulfurization (FGD) to meet an emission level of 520 ng SO_2/J. SRC-II, H-Coal, and EDS liquid fuels can probably meet a 520 ng SO_2/J emission level without FGD.

In order to meet an emission level of 260 ng SO_2/J, (approximately 90 percent control of direct, high sulfur coal-fired emissions), SRC-I would need to have a sulfur content below 0.48 percent. This is well below the typical SRC-I value of 0.7 percent, and would restrict usage of many SRC-I products. SRC-I made from low sulfur coals such as those from Kentucky, Indiana, and the West could meet this level without post-combustion controls, but the use of low sulfur coal negates the purpose of SRC-I production. The heavy liquid fuels from SRC-II, H-Coal, and EDS will probably meet a 260 ng SO_2/J emission level, but in some cases additional hydroprocessing may be required.

To meet an emission level of 90 ng SO_2/J (approximately 96 percent control of direct, high sulfur coal-fired emissions), the use of SRC-I fuel is uncertain (Ref. 2-78). The residual sulfur content of the SRC-I fuel would need to be 0.16 percent. To meet this emission level would require the use of low sulfur coal or flue gas desulfurization processes.

To meet a 90 ng SO_2/J emission level, many of the SRC-II, H-Coal, and EDS liquid fuels will require additional hydroprocessing to reduce their sulfur content to 0.2 percent or less. Table 2.3-10 shows the maximum acceptable sulfur content in the various coal-derived liquid fuels to meet the three emission levels just discussed.

TABLE 2.3-10. EFFECT OF SO_2 EMISSION LEVELS ON SULFUR CONTENT TOLERANCE IN LIQUID FUELS

Fuel Type	Average Heating Value, MJ/kg (Btu/lb)	Typical Sulfur Content of Fuel, wt %	Acceptable Sulfur Content of Fuel, wt % for Flue Gas SO_2 Limits of ng/J (lb/10^6 Btu)		
			520 (1.2)	260 (0.6)	90 (0.2)
SRC-I	34.9 (15,000)	0.7	0.90	0.45	0.15
SRC-I	37.2 (16,000)	0.7	0.96	0.48	0.16
SRC-II, H-Coal, and EDS (raw fuel)	39.5 (17,000)	SRC-II - 0.3 H-Coal - 0.5 EDS - 0.3	1.02	0.51	0.17
SRC-II, H-Coal (hydroprocessed fuels), and EDS	44.2 (19,000)	SRC-II - 0.01 H-Coal - 0.002 EDS - 0.3	1.14	0.57	0.19

B. Nitrogen Oxides--Depending upon the removal of nitrogen from coal in the liquefaction process and/or during hydroprocessing, and the success of developing combustion modifications, flue gas denitrification may or may not be necessary. The results of the SRC-I tests at Plant Mitchell indicate NO_x emissions around 190 ng NO_x/J (Ref. 2-76, 2-77). Unfortunately, these figures are somewhat suspect due to excess air problems experienced while operating the boilers. Further testing is required to determine the actual NO_x content of the combustion gas.

Laboratory testing done on an 880 kW boiler (Ref. 2-79) by KVB, Inc., revealed that by using combustion conditions typical of many large utility boilers, NO_x levels could be reduced to 170 ng NO_x/J using SRC fuel oil (SRCO).

C. Particulate Matter--No conclusive results were obtained from the Plant Mitchell test site for particulate matter. It is known that the ash content of the base coal is reduced to between 0.1 and 0.6 percent in the SRC-I process. It is also thought that burning of the low-sulfur, low-ash SRC will probably produce a high-resistivity fly ash, which may reduce the collection efficiency of conventional cold precipitators which operate near 150°C.

For coal liquids the particulate emissions will depend more on the percent of carbon unburned during combustion than on the amount of ash in the coal liquids. Coal liquids have a very low ash content, and one test on EDS (Ref. 2-63) showed that practically all the carbon is combusted. Table 2.3-11 compares the particulate emissions (Ref. 2-63) from EDS coal liquids with those from low-sulfur fuel oil (LSFO) and regular-sulfur fuel oil (RSFO).

TABLE 2.3-11. EDS FUEL OIL COMBUSTION TEST - PARTICULATE EMISSIONS

Particulate emissions	Coal liquids 205-538°C	Coal liquids 205°C+	Petroleum fuel oils LSFO*	Petroleum fuel oils RSFO*
Total particulates mg/m^3	27	33	50	131
Total particulates, % wt. (in fuel)	0.03	0.04	0.06	0.07
Ash content of sample % wt.	0.027	0.03	0.02	0.08

*LSFO: Low-sulfur fuel oil
RSFO: Regular-sulfur fuel oil

Source: Ref. 2-63

2.3.5.2 Boiler Impacts and Maintenance Requirements--

To date, no commercial coal liquefaction plants have been built and only limited combustion tests have been performed on the coal-derived liquids. As a result, information regarding maintenance requirements and the impact these coal-derived fuels would have on an industrial boiler are not available. However, the impacts and maintenance requirements for coal liquids-fired boilers should be similar to those of oil-fired boilers.

References

2-1. Environmental Protection Agency. Compilation of Air Pollutant Emission Factors, 2nd ed. with supplements. Report No. AP-42. Research Triangle Park, NC, 1973.

2-2. Cavanaugh, E.C., W.E. Corbett and G.C. Page. Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology, Volumes I & II. Final Report. EPA 600/7-77-125a and b, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. September 1977.

2-3. McDowell Wellman Engineering Co. "Wellman-Galusha Gas Producers," Form No. 576. Company Brochure. Cleveland, OH. 1976.

2-4. Dravo Corp. Handbook of Gasifiers and Gas Treatment Systems. Final Report. Task Assignment No. 4. Report No. FE-1772, ERDA Contract No. E(49-18)-1772. Pittsburgh, PA. Chemical Plants Div. February 1976.

2-5. El Paso Natural Gas Co. Application of El Paso Natural Gas Co. for a Certificate of Public Convenience and Necessity. Docket No. CP73-131. El Paso, TX. 1973.

2-6. Woodall-Duckham, Ltd. Trials of American Coals in a Lurgi Gasifier at Westfield, Scotland. Final Report. Research & Development Report No. 105, FE-105. Crawley, Sussex, England. November 1974.

2-7. Rawdon, A.H., R.A. Lisauskas and S.A. Johnson. "Operation of a Commercial Size Riley-Morgan Coal Gasifier." Presented at the American Power Conference. Chicago, IL. 19-21 April 1976.

2-8. Rahfuse, R.V., G.B. Goff and A.J. Liberatore. Noncaking Coal Gasified in a Stirred-Bed Producer. Report No. PB 231-983, Clean Energy Program, Tech. Progr. Rept. No. 77. Morgantown, WV. Morgantown Energy Research Center. 1974.

2-9. Rahfuse, R.V., A.J. Liberatore and G.R. Friggens. Gasification of Caking-Type Bituminous Coal at 75 to 150 psig in a Stirred-Bed Gas Producer. Report No. MERC/PPR-75/3. Morgantown, WV. Morgantown Energy Research Center. July 1975.

2-10. University of Pittsburgh, School of Engineering. Second Annual Conference on Coal Gasification and Liquefaction, Pittsburgh, PA, 5-7 August 1975, collection of papers presented.

2-11. U.S. Energy Research & Development Admin. Quarterly Technical Progress Report, April-June 1976, Grand Forks Energy Research Center. Report No. GFERC/QTR-7674. Grand Forks, ND. GFERC. September 1976.

2-12. Gronhovd, G.H. "Pilot-Plant Experiments in Slagging Gasification," Chem. Eng. Progr., Symp. Ser. 61(54), 104-13(1965).

2-13. Lacey, J.A. "The Gasification of Coal in a Slagging Pressure Gasifier," Amer. Chem. Soc., Div. Fuel Chem., Prepr. 10(4), 151-67(1966).

2-14. Farnsworth, J. Frank, D. Michael Mitsak and J.F. Kamody. "Clean Environment with Koppers-Totzek Process" in Symposium Proceedings: Environmental Aspects of Fuel Conversion Technology, St. Louis, MO, May 1974. Report No. EPA-650/2-74-118, EPA Contract No. 68-02-1325, Task 6. Research Triangle Park, NC. Research Triangle Inst., EPA. October 1974. (pp. 115-30)

2-15. Kamody, John F., and J. Frank Farnsworth. "Gas from the Koppers-Totzek Process and Steam and Power Generation." Presented at the Industrial Fuel Conference, Purdue Univ., West Lafayette, IN, October 1974.

2-16. Air Products and Chemicals, Inc. Engineering Study and Technical Evaluation of the Bituminous Coal Research, Inc., Two-Stage Super Pressure Gasification Process. Report No. PB-235 778, OCR R&D Report 60, OCR Contract No. 14-32-0001-1204. Washington, D.C. OCR. February 1971.

2-17. Robson, Fred L., et al. Fuel Gas Environmental Impact: Phase Report. Report No. EPA-600/2-75-078, EPA Contract No. 68-02-1099. East Hartford, CT. United Technologies Research Center. November 1975.

2-18. Ferrell, J., and G. Poe. Impact of Clean Fuels Combustion on Primary Particulate Emissions from Stationary Sources. Report No. PB-253 452, EPA Contract No. 68-02-1318. Mountain View, CA. Acurex Corp., Aerotherm Div. March 1976.

2-19. Kasai, Takeshi. "Konox Process Removes H_2S," Hydrocarbon Process. 54(2), 93-95(1975).

2-20. Richardson, James T. "SNG Catalyst Technology," Hydrocarbon Process. 52(12), 91-95(1973).

2-21. Bratzler, K., and A. Doerger. "Amisol Process Purifies Gases," Hydrocarbon Processing 53(4), 78-80(1974).

2-22. Dingman, J.C., and T. F. Moore. "Compare DGA and MEA Sweetening Methods," Hydrocarbon Process. 47(7), 138-40(1968).

2-23. Riesenfeld, F.C., and A.C. Kohl. Gas Purification, Second Edition. Houston, TX. Gulf Publishing Co. 1974.

2-24. Rosenwald, R.H., et al. "Sulfide Oxidation: Sulfox System of Pollution Control," Ind. Eng. Chem., Prod. Res, Div. 14(4), 218-26(1975).

2-25. Sulfur Recovery Process Sweetens Sour Gas," Chem. Eng. News 48(18), 68-69(1970).

2-26. Raney, Donald R. "Remove Carbon Dioxide with Selexol," Hydrocarbon Process. 55(4), 73-75(1976).

2-27. Talley, Dennis B. "Startup of a Sour Gas Plant," Hydrocarbon Process. 55(4), 90-92(1976).

2-28. Reid, L.S., and F.M. Townsend. "New Two-Way Process Sweetens Gas and Recovers Sulfur," Oil Gas J. 56(41), 120-22, 124(1958).

2-29. Maddox, R.N. Gas and Liquid Sweetening. Norman, OK. John M. Campbell Co. 1974.

2-30. Woertz, R.B. "A Selective Solvent for Purifying Natural Gas," J. Petrol. Tech. 23 483-90 (April 1971).

2-31. Booz-Allen Applied Research. Evaluation of Techniques to Remove and Recover Sulfur Present in Fuel Gases Produced in Heavy Fossil Fuel Conversion Plants. Report No. 9075-015, EPA Contract No. 68-02-1358. Bethesda, MD. January 1975.

2-32. Goar, B.G. "Sulfinol Process Has Several Key Advantages," Oil Gas J. 67 117-20 (30 June 1969).

2-33. Grekel, H. "H_2S to S...by Direct Oxidation," Oil Gas J. 57(30), 76 (1959).

2-34. "Hydrocarbon Processing Gas Processing Handbook," Hydrocarbon Process. 54(4), 79-138(1975).

2-35. Doerges, A., K. Bratzler and J. Schlauer. "LUCAS Process Cleans Lean H_2S Streams," Hydrocarbon Process. 55(10), 110-11(1976).

2-36. Rawdon, A.H., R.A. Lisauskas and S.A. Johnson. "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames," Presented at the NO_x Control Technology Seminar, Sponsored by Electric Power Research Institute, San Francisco, CA, 5-6 February 1976.

2-37. Spaite, Paul W., and Gordon C. Page. Low- and Medium-Btu Gasification Systems: Technology Overview. Final Report. Report No. EPA-600/7-78-061, EPA Contract Nos. 68-02-2149 and 68-02-2147. Cincinnati, OH, and Austin, TX. Paul W. Spaite Co. and Radian Corp. March 1978.

2-38. Dressel, Jack H., Joseph H. Asbil. "Prepare Coal for Gasification," Hydrocarbon Process. 53(3), 91-93(1974).

2-39. Forney, A.J., and J.P. McGee. "The Synthane Process - Research Results and Prototype Plant Design," Presented at the Fourth AGA Synthetic Pipeline Gas Symposium, Chicago, IL, October 1972.

2-40. Applying Air Pollution Control Equipment. Environmental Handbook Series. Reprinted from Pollution Engineering.

2-41. U.S. Bureau of Reclamation, Upper Colorado Region. WESCO Gasification Project and Expansion of Navajo Mine by Utah International Inc., San Juan County, New Mexico, Draft Environmental Statement. DES 74-107. Washington, DC. 1974.

2-42. Jahnig, C.E. Evaluation of Pollution Control in Fossil Fuel Conversion Processes. Gasification, Section 5: Bi-Gas Process. Final Report. Report No. PB-243 694, EPA 650/2-74-009g, EPA Contract No. 68-02-0629. Linden, NY. Exxon Research & Engineering Co. May 1975.

2-43. Jahnig, C.E. Evaluation of Pollution Control in Fossil Fuel Conversion Processes. Gasification, Section 8: Winkler Process. Final Report. Report No. EPA-650/2-74-009j, EPA Contract No. 68-02-0629. Linden, NJ, Exxon Research & Engineering Co. September 1975.

2-44. Jensen, G.A. "The Kinetics of Gasification of Carbon Contained in Coal Minerals," Research Results Ms. 74-126. Washington, DC. ACS. 1974.

2-45. Coles, W.F., and J.R. Stewart. "Considerations When Converting Industrial Plants to Coal Firing," Presented at the ASME Industrial Power Conference, Cleveland, OH, October 1977.

2-46. Babcock and Wilcox. Fossil Fuel and Advanced Systems. Final Report. PB-251 442/OST, EPRI-265-2. Palo Alto, CA. EPRI. January 1976.

2-47. Southern Research Inst. A Manual of Electrostatic Precipitator Technology, Pt. 2, Application Areas. Contract No. CPA 22-69-73. Birmingham, AL. 1970.

2-48. Thomas, W.C., K.N. Trede, and G.C. Page. Environmental Assessment: Source Test and Evaluation Report - Wellman-Galusha (Glen Gery) Low-Btu Gasification. Final Report. EPA Report No. 600/7-79-185, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. August, 1979.

2-49. Page, Gordon C. Environmental Assessment: Source Test Evaluation Report--Chapman Low-Btu Gasification. Final Report. EPA-600/7-78-202. EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. October 1978.

2-50. Waibel, R.T., E.S. Fleming, and D.H. Larson. Pollutant Emissions From "Dirty" Low- and Medium-Btu Gases. Final Report. EPA-600/7-78-191. EPA Contract No. 68-02-2643. Chicago, IL. Institute of Gas Technology. October 1978.

2-51. Telephone communications between P.J. Murin (Radian Corp.) and Tom Moore (Jefferson Chemical) concerning performance of DGA Acid Gas Removal Process. 2 June, 13 July, and 18 July 1978.

2-52. "NG/SNG Handbook," Hydrocarbon Process. 50(4), 93-122(1971).

2-53. Perry, Charles R. "Basic Design and Cost Data on MEA Treating Units," in Proceedings of the 1967 Gas Conditioning Conference, University of Oklahoma, Norman, OK.

2-54. Radian Corporation. Environmental Control Selection Methodology for a Coal Conversion Demonstration Facility. Final Report. DOE Contract No. EX-76-01-2314. Austin, TX. February 1978.

2-55. Fossil Energy Research and Development Program of the U.S. Department of Energy. DOE/ET-0013 (78), U.S. Department of Energy. 1975.

2-56. Schmid, B.I. and D.M. Jackson. Recycle SRC Processing for Liquid and Solid Fuels. In: Paper presented at the Fourth Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity. August 2-4, 1977. Pittsburgh, PA.

2-57. "Demonstration Plants Set for SRC Process," Oil & Gas Journal 76(30), 24(1978).

2-58. Hittman Associates, Inc. Environmental Assessment Data Base for Coal Liquefaction Technology, Vol. I & II, Final Report. EPA-600/7-78-184a & b, EPA Contract No. 68-02-2162. Columbia, MD. September 1978.

2-59. U.S. Environmental Protection Agency. Task 7 - Summary Report: Technical and Economic Bases for Evaluation of Emission Reduction Technology. EPA Contract No. 68-02-2603. Prepared by PEDCo Environmental, Inc., Cincinnati. 1978.

2-60. Babcock & Wilcox. Steam: Its Generation and Use. New York, New York. 1975.

2-61. Callen, Robert B., et al. Upgrading Coal Liquids to Gas Turbine Fuels. 1, Analytical Characterization of Coal Liquids, I&EC Product Research and Development, 15(4):222-233. 1976.

2-62. Cabal, A.V., et al. Utilization of Coal-Derived Liquid Fuels in a Combustion Turbine Engine. In: Preprint of Papers Presented at the American Chemical Society Symposium at Anaheim, California. 23(1). March 1978.

2-63. Quinlan, C.W. and C.W. Siegmund. Combustion Properties of Coal Liquids From the Exxon Donor Solvent Process. In: Preprint of Papers Presented at the American Chemical Society Symposium at Anaheim, California, 23(1). March 1978.

2-64. Muzio, L.J. and J.K. Araad. Small Scale Evaluation of the Combustion and Emission Characteristics of SRC Oil, In: Preprints of Papers Presented at the American Chemical Society Symposium at Anaheim, California, 23(1). March 1978.

2-65. Sage, W.L., W. Downs and P.L. Cioffi. Characteristics of Solvent Refined Coal: Dual Register Burner Tests, FP-628, EPRI 1235-5. Prepared for Electric Power Research Institute by the Babcock & Wilcox Company, Ohio. 1978.

2-66. Borio, R.W., et al. Solvent Refined Coal Evaluation: Pulverization, Storage, and Combustion. EPRI 1235-2b. Prepared for Electric Power Research Institute by C-E Power Systems. Connecticut. June 1976.

2-67. Downs, W., et al. Investigating Storage, Handling and Combustion Characteristics of Solvent Refined Coal: Task C Extension, EPRI 1235-4. Prepared for Electric Power Research Institute by the Babcock & Wilcox Company. 1976.

2-68. Borio, R.W. Laboratory Analysis of Solvent Refined Coal. EPRI 1235-2a. C-E Power Systems. Combustion Engineering, Inc. June 1976.

2-69. U.S. Department of Energy. H-Coal Integrated Pilot Plant: Phase I - Final Report, Contract No. EX-76-01-1544. Prepared for U.S. Department of Energy by Hydrocarbon Research Inc. November 1977.

2-70. Fant, B.T. Exxon Donor Solvent Coal Liquefaction Commercial Plant Study Design. Prepared for U.S. Energy Research and Development Administration by Exxon Research and Engineering Company. Florham Park, New Jersey. January.

2-71. Balzhiser, R.E. R&D Status Report: Fossil Fuel and Advanced Systems Division, EPRI Journal, Vol. 3, No. 5. June 1978.

2-72. Nongbri, G. and N. Stewart. Solvent Refining of West Kentucky 9-14 Coal. EPRI Af-499. Prepared for Electric Power Research Institute by Mobil Research and Development Corporation. Paulsboro, NJ. October 1977.

2-73. Dabkowski, M.J., et al. Upgrading of Coal Liquids for Use as Power Generation Fuels. EPRI AF-44. Prepared for Electric Power Research Institute by Mobil Research and Development Corporation. Paulsboro, NJ. October 1977.

2-74. Epperly, W.R. EDS Coal Liquefaction Process Development. Phase IIIB. FE-2893-3. Prepared for U.S. Department of Energy by Exxon Research and Engineering Company. Florham Park, NJ. July-September 1977.

2-75. Spencer, D.F., S.B. Alpert and R.H. Wolk. Review of Alternative Liquefaction Processes. EPRI: Electric Power REsearch Institute, Palo Alto, CA. 1978.

2-76. The Southern Company Services, Solvent Refined Coal Burn Test. April 1978.

2-77. Budden, Kenneth G., and Subhash S. Patel. Air Emissions from Combustion of Solvent Refined Coal. Final Report. Report No. EPA-600/7-79-004, EPA Contract No. 68-02-2162. Columbia, MD, Hittman Associates, Inc., January 1979.

2-78. Alpert, S.B., et al. Review of Solvent-Refined Coal Technology. Electric Power Research Institute. March 1978.

2-79. Arand, J.K. and L.J. Muzio. Combustion and Emission Evaluation of SRC Fuel Oil, A Synthetic Liquid from Coal, KVB 19900-733. Prepared for Gulf Mineral Resources by KVB, Inc. Tustin, CA. December 1977.

SECTION III

SELECTION OF "BEST CANDIDATE" SYNTHETIC FUELS SYSTEMS

In the previous section, the population of coal gasification and liquefaction systems were identified and discussed. From that population, a list of systems considered to be likely candidates for use as precombustion emission controls for industrial boilers were identified. For the coal liquefaction systems, the preliminary screening step identified four candidate systems: SRC-I, SRC-II, H-Coal, and Exxon Donor Solvent (EDS). For low- and medium-Btu coal gasification, fourteen candidate gasifiers and fourteen acid gas removal (AGR) processes were identified (see Tables 2.1-6 and 2.1-7).

The second step in assessing the synfuels from coal technologies is to compare the candidate systems and select several "best candidate" systems for detailed analyses of costs and energy and environmental impacts (see Figure 3.1-1). The results of those comparisons and selections are presented in this section. Low-Btu coal gasification is addressed in Section 3.1, medium-Btu gasification in Section 3.2, and coal liquefaction in Section 3.2.

In selecting "best candidate" systems, the following criteria were considered:

- System performance,
- Status of development,
- Applicability,

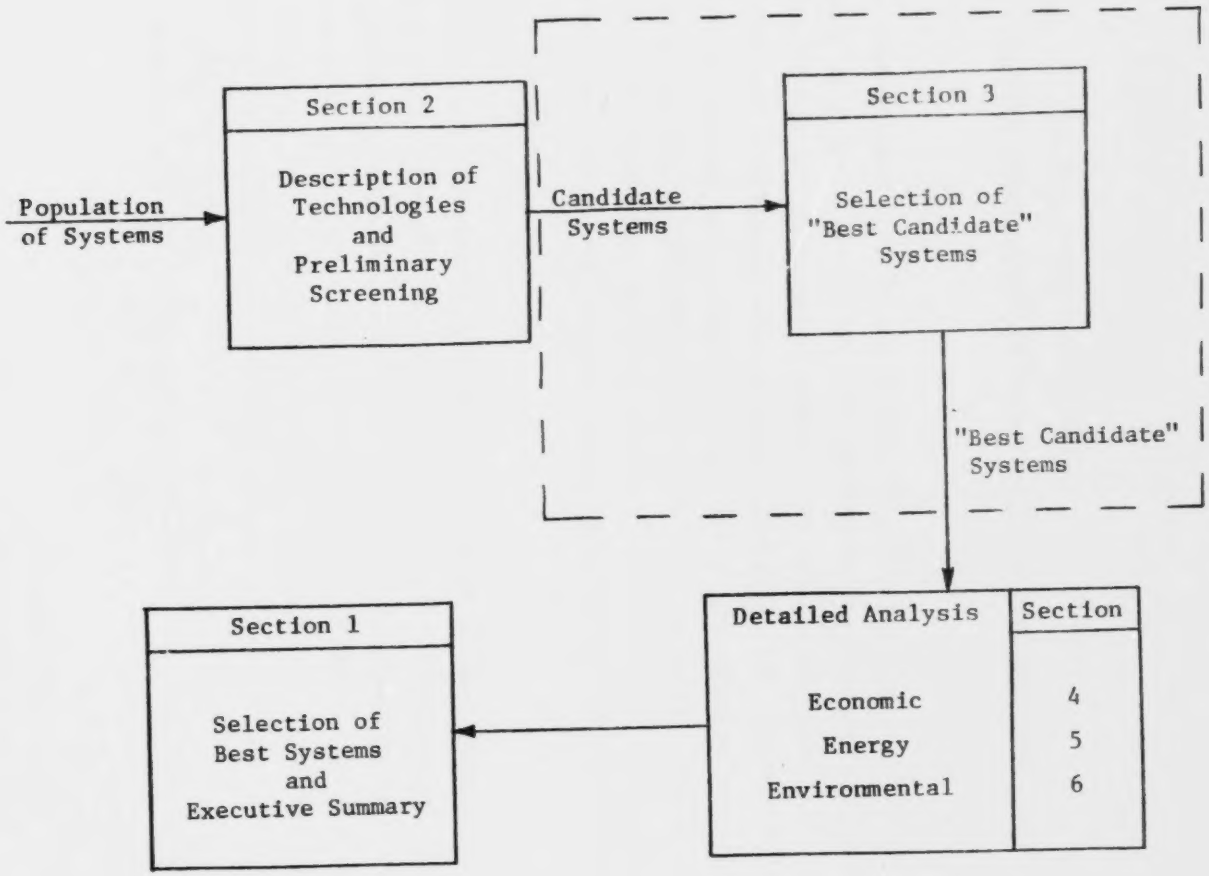

Population of Systems
Section 2
Description of Technologies and Preliminary Screening
Candidate Systems
Section 3
Selection of "Best Candidate" Systems
"Best Candidate" Systems
Detailed Analysis
Economic
Energy
Environmental
Section
4
5
6
Section 1
Selection of Best Systems and Executive Summary

Figure 3.1-1. Schematic of study approach.

- Preliminary cost considerations,
- Preliminary energy considerations, and
- Preliminary environmental considerations.

A discussion of how these criteria were applied to the synthetic fuels from coal systems is presented in the following text.

Obviously, a primary concern in the selection of a synthetic fuels system is the system's performance, i.e., what emissions will result from combustion of the synthetic fuel. In applying this selection criteria, system performance was compared to three target control levels for SO_x, NO_x, and particulate emissions. Labeled "moderate, intermediate, and stringent", these target control levels were chosen only to encompass all candidate technologies and form bases for comparison of technologies for control of specific pollutants considering performance, costs, energy, and non-air environmental effects. Table 3.1-1 summarizes the target emission control levels selected for the synthetic fuels technologies. The following text briefly discusses the rationale behind the selections.

TABLE 3.1-1. TARGET EMISSION CONTROL LEVELS FOR SYNTHETIC FUELS FROM COAL TECHNOLOGIES

	SO_2, ng/J (lb/10^6 Btu)	NO_x ng (as NO_2)/J (lb/10^6 Btu)	Particulates, ng/J (lb/10^6 Btu)
Coal Gasification			
Moderate Control	150 (0.5)	86 (0.2)	13 (0.03)
Intermediate Control	86 (0.2)		
Stringent Control	43 (0.1)		
Coal Liquefaction			
Moderate Control	520 (1.2)	300 (0.7)	13 (0.03)
Intermediate Control	260 (0.6)	220 (0.5)	
Stringent Control	86 (0.2)	86 (0.2)	

The moderate SO_2 target control level for gasification systems was set at 150 ng/J (0.35 lb/10^6 Btu) heat input to the boiler. This emission level is what would be obtainable by removing essentially all of the H_2S from a high sulfur coal derived low-Btu gas. The stringent SO_2 control level was set at 43 ng/J (0.1 lb/10^6 Btu). The NO_x and particulate emission target control levels were set at 86 ng NO_x/J (0.2 lb/10^6 Btu) and 13 ng particulates/J (0.03 lb/10^6 Btu). These levels were selected because they are comparable to those for natural gas combustion. Also, because these levels are considered low, but should be readily attainable by this technology, only one target level of control was chosen for NO_x and particulates.

For the coal liquefaction systems, the moderate target SO_2 control level was set at 520 ng/J (1.2 lb/10^6 Btu) heat input to the boiler. This level is what appears to be attainable by the SRC-I process (the most advanced liquefaction system) using a high sulfur coal feed. The stringent target control level was set at 86 ng/J (0.2 lb/10^6 Btu), which represents what would be obtainable by additional hydroprocessing of the synthetic liquids produced by the SRC-II, H-Coal, or EDS processes. The moderate target NO_x control level was set at 300 ng/J (0.7 lb/10^6 Btu), which is what would be expected from direct coal combustion. This is a reflection of the difficulty that the liquefaction processes have in removing nitrogen from the synthetic liquids. The stringent target NO_x control level was set at 86 ng/J (0.2 lb/10^6 Btu). This level is equal to the gasification target control level and could be obtainable with additional (possibly extensive) hydroprocessing. Only one target particulate control was selected - 13 ng/J (0.03 lb/10^6 Btu). This level is what would normally be expected from combustion of low ash containing synthetic liquid fuels.

In choosing the "best candidate" systems for achieving a given emission control level, system performance is evaluated for two coal feedstocks--a high sulfur eastern coal and a low sulfur western coal (see Table 3.1-2). Examining other coals, such as a low or medium sulfur eastern or medium sulfur western, was not felt to be necessary. The rationale for this decision can best be explained by considering, for example, low-Btu coal gasification. By including sulfur species removal in the low-Btu gasification system, several constraints are imposed on the processing scheme. Specifically, the currently available sulfur species removal techniques require the feed gas to be at a relatively low temperature, and essentially free of particulate matter. Thus, prior to sulfur removal the raw gas from the gasifier is treated in a series of quenching and cooling units. A side benefit of the quenching/cooling steps is the removal of particulate matter and most of the nitrogen bearing compounds (NH_3 and HCN) present in the raw low-Btu gas. The amounts of particulates and nitrogen species removed have little influence on the energy, environmental, and cost impacts of the gasification system. Therefore, control of particulate and NO_x emissions via low-Btu coal gasification has minimal dependence on the coal feedstock composition.

As would be expected though, the impacts of controlling sulfur emissions via coal gasification do depend on the coal feed sulfur content. However, the quantity of sulfur species present in the raw low-Btu gas (and hence the quantity which must be removed to meet a given SO_2 emission control level) is directly proportional to the coal sulfur content. Thus, by examining both a low and high sulfur coal feed, the range of impacts will be examined. And, from that information the impacts associated with a medium sulfur coal could be easily interpolated.

Status of development is also an important selection criteria. Emission standards for industrial boilers are expected to be proposed by 1981. Therefore, the systems being assessed in this report must be either currently commercially available or have reasonable expectations of being successfully demonstrated and commercially available by 1981.

TABLE 3.1-2. ULTIMATE ANALYSIS OF BASE COALS

	Eastern High Sulfur	Western Low Sulfur
Composition, wt %		
Carbon	64.80	57.60
Hydrogen	4.43	3.20
Sulfur	3.54	0.60
Oxygen	6.56	11.20
Nitrogen	1.30	1.20
Moisture	8.79	20.80
Ash	10.58	5.40
Heating Value, MJ/kg (Btu/lb)	27.4 (11,800)	22.3 (9,600)

Applicability is the third major criteria used in selecting "best candidate" systems. Some candidate control systems may not be feasible for certain boiler types or capacities.

Costs, energy, and environmental impacts of the emission control systems are as important as the three criteria mentioned above. However, for the process screening step discussed in this section, only preliminary estimates of these factors are used. Detailed analysis of costs, energy, and environmental impacts will be performed in Sections 4, 5 and 6 respectively for each of the "best candidate" systems selected.

3.1 SELECTION OF "BEST CANDIDATE" LOW-BTU COAL GASIFICATION SYSTEMS

In selecting "best candidate" low-Btu gasification systems, it is necessary to select not only "best candidate" gasifiers, but more importantly to select "best candidate" acid gas removal (AGR) processes. As discussed in Section 2, the performance of a gasification system as a precombustion emission control depends on its ability to produce a "clean" burning fuel. Sulfur compounds are removed by the AGR process employed, while particulates and fuel bound nitrogen compounds (e.g., NH_3 and cyanides) are mainly removed when the raw low-Btu gas is quenched and cooled. Since quenching/cooling is a prerequisite for all of the AGR processes considered in this study, control of particulate and NO_x emissions is a side benefit of SO_2 emissions control.

The candidate gasifiers identified in Section 2 are compared in Section 3.1.1 and a "best" or representative candidate is selected. The compositions of the raw and cooled low-Btu gases produced by this gasifier from the two base case coal feedstocks are also presented in this section. Then, in Section 3.1.2 the candidate AGR processes are compared and "best candidate" processes selected. Finally, Section 3.1.3 is a brief summary of the "best candidate" systems selected, i.e., the combination of gasifiers and acid gas removal processes for which detailed analyses will be performed in Sections 4, 5 and 6.

3.1.1 Comparison and Selection of a Candidate Low-Btu Gasifier

Fourteen candidate low- and/or medium-Btu gasifiers were identified in Section 2. Examination of this list indicates that four of those gasifiers can only be operated with pure (98+ percent) oxygen as the gasifying agent and hence are considered as medium-Btu gasifiers. The ten gasifiers which can use air as the gasifying agent, as well as the four medium-Btu gasifiers, are listed in Table 3.1-3. Also included in this table are the data which were used to compare the low-Btu gasifiers. The result of the comparisons was the selection of the Wellman-Galusha gasifier as the "best candidate". The rationale behind that selection is presented in the following text.

Six of the ten low-Btu gasifiers listed in Table 3.1-3 are commercially available. In addition, the Foster Wheeler/Stoic, which is currently being demonstrated, is nearly identical to the commercially available Woodall-Duckham/Gas Integrale gasifier and as such is considered commercially available. The other three gasifiers, BI-GAS, Coalex, and Pressurized Wellman Galusha (MERC) are in the pilot or demonstration plant stage of

TABLE 3.1-3. COMPARISON OF CANDIDATE GASIFIERS

Gasifier	Type of Gas	Pressure	Development Status	Feedstock Restrictions	Environmental Impacts		Cold Gas Efficiency (%)[a]
					Direct	Indirect	
BGC Slagging Lurgi	Medium-Btu	High Pressure	Demonstration plant (started 1976)	Accepts all types; strongly caking requires agitation; 13-51 mm size coal; <20% moisture; crushing and sizing required; low ash coal may require fluxing agents.	Coal lock gas Slag quench vent gas Slag lock gas Slag quench blowdown Slag slurry	Waste streams from gas purification; process condensate quenching liquor, coal fines.	83
Bi-Gas	Low- or Medium-Btu	High Pressure	Not been demonstrated; pilot plant started in August 1976	Accepts all coal types; 70% less than 0.074 mm; crushing and pulverizing required; also slurry preparation or fluxing agent.	Slurry preparation vent gas Slag lock gas Slag slurry (solid & liquid)	Essentially no tars, oils, naphthas, or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	69
Chapman (Wilputte)	Low- or Medium-Btu	Atmospheric	Commercially available (since 1945) for Low-Btu gas; not commercially demonstrated for medium-Btu gas	Accepts all coal types; size <102 mm; crushing and sizing required.	Barrel valve vent gas Poke hole gases Ash pan gas Ash	Waste streams from gas purification; process condensate, quenching liquor; coal fines.	DNA
Coalex	Low-Btu	Atmospheric	Pilot plant since 1976	Accepts all coal types (lignite not been tested); <0.07 mm sized coal; crushing and pulverizing required; additives required.	Additive hopper vent gas Slag slurry (solid & liquid)	Essentially no tars or oils in raw product gas; gas purification may create additional unwanted streams.	DNA
Foster Wheeler/ Stoic	Low-Btu	Atmospheric	Demonstration plant construction begun in 1977	Lignite, subbituminous, noncaking bituminous; 19-38 mm sized coal; crushing and sizing required; partial oxidation may be needed for strongly caking coals.	Coal hopper vent gas Ash pan gas Ash Poke hole gases	Waste streams from gas purification; tars/oils, coal dust.	77
GFERC Slagging	Medium-Btu	High Pressure	Pilot plant (1958-1965)	Bituminous char, lignite char, lignite; 6.4-19 mm sized coal; <35% moisture; Crushing and sizing required.	Coal lock gas Slag quench vent gas Slag quench blowdown Slag slurry	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	85
Koppers-Totzek	Medium-Btu	Atmospheric	Commercially available since 1952	Accepts all coal types; 70% to 90% less than 0.074 mm; pulverizing required; ~1-8% moisture; possible addition of fluxing agents to lower ash fusion temp.	Coal bin N_2 vent Slag Slag slurry	Essentially no tars, oil, naphthas or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	75

(Continued)

TABLE 3.1-3. Continued

Gasifier	Type of Gas	Pressure	Development Status	Feedstock Restrictions	Environmental Impacts		Cold Gas Efficiency (%)[a]
					Direct	Indirect	
Lurgi	Low- or Medium-Btu	High Pressure	Commercially available (since 1941)	Accepts all coal; strongly caking coals may need partial oxidation or agitation; crushing and sizing required; <35% moisture; 3.2-38.1 mm sized coal.	Coal lock gas Ash lock gas Ash Ash quench water	Waste streams from gas purification; process condensate, quenching liquor, tars, oil, phenols, NH_3.	63-80
Pressurized Wellman-Galusha (MERC)	Low- or Medium-Btu	High Pressure	Pilot plant (since 1958); not commercially demonstrated for medium-Btu	Accepts all types; 50% <12.7 mm sized; no predrying; crushing and sizing required.	Coal lock gas Ash lock gas Ash	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	79
Riley-Morgan	Low- or Medium-Btu	Atmospheric	Commercially available for low-Btu gas (Pilot plant since 1975); not commercially demonstrated for medium-Btu gas	Accepts all coal types; 3.2-51 mm sized coal; crushing and sizing required.	Coal lock gas Ash pan gas Ash Poke hole gases	Waste streams from gas purification; process condensate, quenching liquor, coal fines.	64-68
Texaco	Medium-Btu	High Pressure	Pilot plant	Accepts all coal types; 70% less than 0.074 mm; crushing, pulverizing, slurry preparation required.	Slurry preparation vent gas Slurry steam purge Preheater flue gases Slag slurry	Essentially no tars, oils, naphthas, or phenols in raw product gas; gas treatment produces process condensate and quenching liquor.	77
Wellman-Galusha	Low-Btu	Atmospheric	Commercially available (since 1941)	Can use anthracite, bituminous, charcoal, or coke; 7.9-14.3 mm for anthracite; 26-51 mm for bituminous; crushing and sizing required.	Coal bin gas Ash hopper gas Ash Poke hole gases	Waste streams from gas purification; coal fines, process condensate, quenching liquor.	75
Winkler	Low- or Medium-Btu	Atmospheric	Commercially available since 1926	Lignite, subbituminous, weakly caking bituminous; <9.53 mm sized coal; crushing required; <30% moisture for lignites; <18% moisture for higher rank coals; partial oxidation may be required.	Coal bin N_2 vent Dry ash bin N_2 vent Ash slurry settler vent Dry ash Ash slurry	Essentially no tars, oils, or naphthas in raw product gas; treatment produces process condensate and quenching liquor.	55-72
Woodall-Duckham/Gas Integrale	Low-Btu	Atmospheric	Commercially available (since 1940)	Lignite, bituminous; 6.4-38.1 mm sized coal; crushing and sizing required; drying not required; partial oxidation for strongly caking coals.	Coal hopper vent gas Ash lock gas Ash	Waste streams from gas purification; tar, process condensate, dust.	77

[a] Cold Gas Efficiency = $\frac{\text{Product Gas Energy Output}}{\text{Coal Energy Input}}$

Source: Ref. 3-1 through 3-16

DNA: Data not available

development. Current plans for these gasifiers do not indicate that they will be commercially available within the time frame being considered in this study. Therefore, they will not be considered further.

Of the seven gasifiers considered commercially available, five are small capacity (coal feed rates approximately 5000 kg/hr or less), atmospheric pressure units (Ref. 3-1). These include:

- Wellman-Galusha,
- Woodall-Duckham/Gas Integrale,
- Chapman,
- Riley-Morgan, and
- Foster Wheeler/Stoic.

The other two gasifiers typically have larger capacities. The atmospheric pressure Winkler unit gasifies around 20,000 kg/hr (44,000 lb/hr), while a 3 m (10 ft) diameter pressurized Lurgi gasifier has a coal feed rate of 5,000-18,000 kg/hr (11,000-40,000 lb/hr) depending on coal type (Ref. 3-1).

For the industrial boiler sizes being considered in this study (10-60 MW), the coal feed rate to a companion sized low-Btu gasification facility would be 2,000-11,000 kg/hr (4,400-24,000 lb/hr). Thus, a single Winkler or Lurgi gasifier would be too large for even the largest capacity boiler being examined, let alone for the smaller boilers. The other gasifiers have capacities comparable to the smaller boilers, and for the larger boilers could meet the fuel demand with the installation of multiple gasifiers.

In light of the status of development and capacity considerations of the ten candidate low-Btu gasifiers, only the five gasifiers listed above are considered as possible "best candidates". Since all five gasifiers also operate at atmospheric pressure, have comparable energy conversion efficiencies, and are relatively simple pieces of equipment, their energy and economic impacts are also comparable. Moreover, the major environmental impacts associated with a gasification system originate in the gas cleanup steps. Thus, based on preliminary estimates of costs, environmental, and energy impacts, none of the five possible "best candidate" gasifiers are eliminated or singled out as being superior. Therefore, since all of these gasifiers are comparable, the Wellman-Galusha gasifier was selected as the representative "best candidate" gasifier because there are more data available concerning its waste streams and product gases produced from various coal types.

The estimated compositions of the raw and quenched/cooled low-Btu gases produced by gasifying the two study basis coals (see Table 3.1-2) in a Wellman-Galush gasifier are shown in Table 3.1-4. The major gaseous components were taken from a study performed by Battelle (Ref. 3-17). The NH_3, HCN, H_2S and COS values for the low sulfur coal case were estimated from experimental data (Ref. 3-7, 3-18, 3-19). For the high sulfur coal case, values for these minor components were based on engineering estimates, since data for gasifying high sulfur coal in an atmospheric pressure, fixed-bed gasifier were not available. Tars and oils in the raw low-Btu gases were estimated at 10 percent of the coal feed.

Entrained particulates and tars/oils in the raw low-Btu gases produced by the Wellman-Galusha gasifier are removed by the quench/cooling steps shown in Figure 3.1-2. These steps include a cyclone for removal of particulates, an in-line quench, tray and spray scrubbers for tar and oil removal, and an electrostatic precipitator to remove residual tars, oils and particulates. Greater than 99.5 percent of the tars, oils and particulates can be removed by this quench/cooling system (Ref. 3-20, 3-21, 3-22).

3.1.2 Comparison and Selection of Candidate Acid Gas Removal Processes

The preliminary screening step discussed in Section 2 identified the following candidate acid gas removal (AGR) processes:

TABLE 3.1-4. ESTIMATED COMPOSITIONS OF LOW-BTU GASES FROM WELLMAN-GALUSHA GASIFIER

	Low Sulfur Western Coal Feed		High Sulfur Eastern Coal Feed	
	Raw Gas	Quenched Gas*	Raw Gas	Quenched Gas*
Composition,** vol. %				
CO_2	2.99	3.11	3.04	3.08
CO	25.06	26.11	25.74	26.01
CH_4	2.47	2.57	2.56	2.59
H_2	12.98	13.53	12.73	12.87
H_2O	12.34	8.67	9.62	8.67
N_2	44.02	45.87	45.53	46.01
H_2S	0.10	0.11	0.69	0.70
COS, ppmv	50	54	270	280
NH_3, ppmv	260	270	270	280
HCN, ppmv	78	81	97	98
Tars & Oils, kg/m^3 (@ 43°C, 1 atm)	0.030	0.0002	0.027	0.0001
HHV, MJ/m^3	5.48	5.71	5.56	5.62

*The amounts of H_2S, COS, NH_3, HCN, as well as major gas components, which will be absorbed from the gas during the quenching steps is not known. The values shown assume no gas absorption. Tar and oil values are based on 99.5% removal.

**Major gas components are based on information presented in Ref. 3-17. H_2S, COS, NH_3, and HCN values are based on experimental data (Ref. 3-7, 3-18, and 3-19). Tars and oils in raw gas are estimated at 10% of the coal feed.

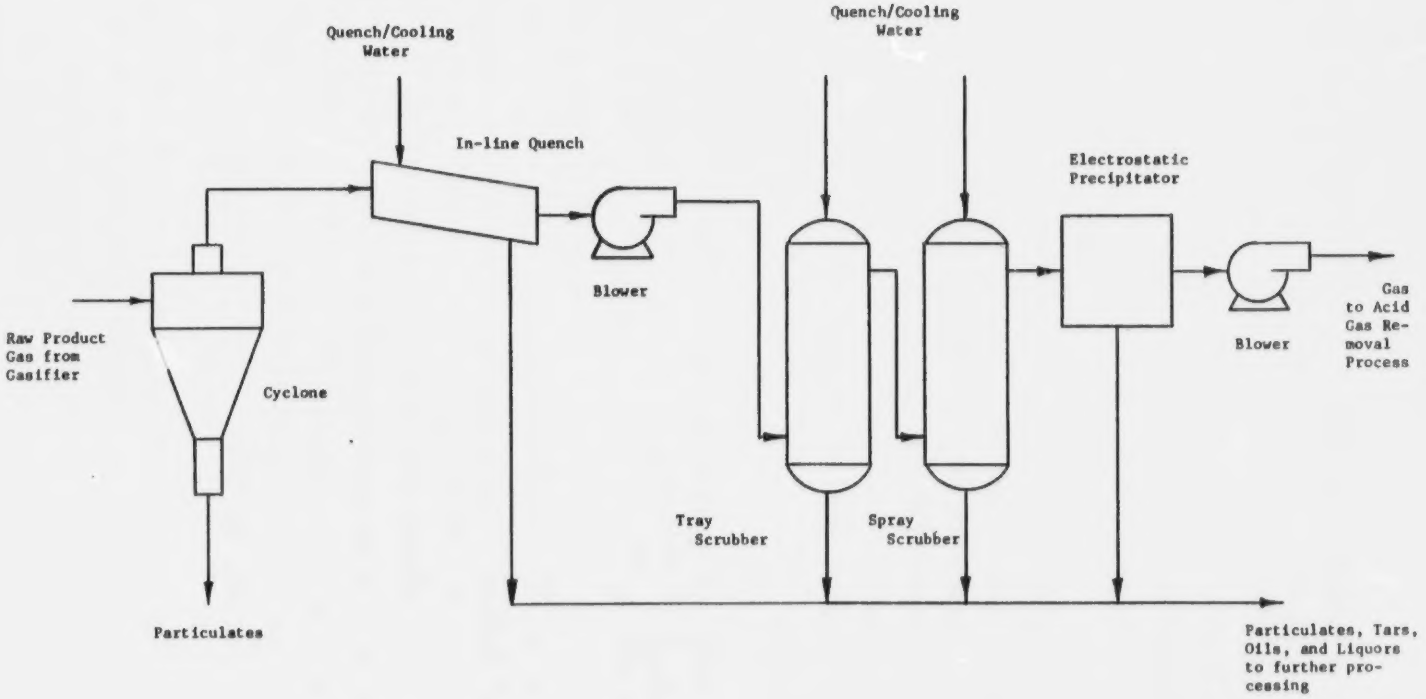

Quench/Cooling
Water
In-line Quench
Quench/Cooling
Water
Electrostatic
Precipitator
Raw Product
Gas from
Gasifier
Cyclone
Blower
Tray
Scrubber
Spray
Scrubber
Blower
Gas
to Acid
Gas Re-
moval
Process
Particulates
Particulates, Tars,
Oils, and Liquors
to further pro-
cessing

Figure 3.1-2. Wellman-Galusha quench/cooling steps.

- Physical Solvent Processes
    - Rectisol
    - Selexol
    - Purisol
    - Estasolvan
    - Fluor Solvent

- Chemical Solvent Processes
    - MEA (monoethanolamine)
    - MDEA (methyldiethanolamine)
    - DEA (diethanolamine)
    - DIPA (diisopropanolamine)
    - DGA (diglycolamine)
    - Benfield

- Combination Chemical/Physical Solvent Processes
    - Amisol
    - Sulfinol

- Direct Conversion Process
    - Stretford

Pertinent data concerning these processes are presented in Tables 3.1-5, 3.1-6, and 3.1-7. Comparisons of these processes with respect to their use in low-Btu gasification facilities are summarized in the following text. The results of the comparisons are the selection of the Stretford and MEA processes as the "best candidates" for further study.

First, all of the candidate processes are commercially available. However, most of the processes have not found application in coal gasification facilities. The most common applications have been in the purification of natural gas or refinery gas. These are high pressure applications that required essentially complete ($\leq$5 ppmv) removal of sulfur species. In spite

TABLE 3.1-5. PHYSICAL SOLVENT ACID GAS REMOVAL SYSTEM

	Estasolvan	Fluor Solvent	Purisol	Rectisol	Selexol
Commercial Application, Purification	Natural gas and liquid hydrocarbon recovery in natural gas	Natural gas, ammonia and hydrogen gas	Natural gas and hydrogen streams	Hydrogen synthesis gas, partial oxidation effluent gas and low/medium Btu coal gas	Synthesis and natural gas, and coal gasification pilot plant product gas
Applicability to Gasification	Technically feasible	Technically feasible	Technically feasible	Proven for pressurized low/medium-Btu coal gas	Technically feasible
Operating Pressure	Typically 6.9 MPa	5.86-6.90 MPa	Typically 6.9 MPa	2.06-6.90 MPa	3.45-6.90 MPa
Maximum Control Effectiveness					
High Pressure H_2S	<3 ppmv	<4 ppmv	2 ppmv	1 ppmv	1 ppmv
High Pressure COS/CS_2	<0.25 g/100 scf	Data not available	Data not available	1 ppmv	1 ppmv
High Pressure R-SH	<2.0 g/100 scf	Data not available	Data not available	1 ppmv	1 ppmv
Low Pressure H_2S, COS/CS_2, R-SH	Ineffective	Ineffective	Ineffective	Ineffective	Ineffective
Operational, Maintenance Problems, and Reliability	Not complex.	Requires low heavy hydrocarbon content and high combined H_2S and CO_2 partial pressure (275 psi). Not complex.	Expensive solvent. Moderately complex operation.	Inexpensive solvent; sustained cooling is required resulting in a complex flow pattern for minimization of energy requirements by use of heat exchangers. This results in moderately difficult start-up and shutdown.	Expensive solvent. Not effective at low acid gas concentrations. Moderately complex.
Environmental Considerations:					
System produces an acid gas stream	Yes	Yes	Yes	Yes	Yes
System produces a solvent blowdown	Yes	Yes	Yes	Yes	Yes
Preliminary Capital Cost	Data not available	Data not available	Data not available	Data not available	Approximately 70 percent of MEA
Utility Requirements:					
Steam	Yes	Yes	Yes	Yes	Yes
Electricity	Yes	Yes	Yes	Yes	Yes
Cooling Water	Yes	Yes	Yes	Yes	Yes

Source: Ref. 3-1, 3-2, 3-23 through 3-26

TABLE 3.1-6. CHEMICAL SOLVENT ACID GAS REMOVAL SYSTEMS

		BENFIELD	DEA	DGA	DIFA	MDEA	MEA
Commercial Application, Purification		Pilot plant gasification producer gas, sour natural gas	Natural gas and refinery raw gas	Natural, synthesis, and refinery sour gases	Natural gas, refinery gas, synthesis gas or LPG	Refinery sour gases	Refinery gas
Applicability to Gasification		Proven for low/medium Btu gas	Technically feasible	Technically feasible (applied on pilot plant)	Technically feasible	Technically feasible	Technically feasible
Operating Pressure		0.69-3.18 MPa	atm +	Not pressure-sensitive (atm +)	0.1-6.90 MPa	0.10-0.69 MPa	Not highly pressure-sensitive (atm +)
Maximum Control Effectiveness							
High Pressure	H_2S	2 ppmv	3 ppm H_2S	<4 ppmv	5 ppmv (@ 6.9 MPa)	4 ppmv	<1 ppmv
	COS/CS_2	27 ppm	70	90+	Some	Most	95+
	R-SH	Data not available	Some	- -	Some	Data not available	- -
Low Pressure	H_2S	Data not available		Data not available	Less than 100 ppmv	4 ppmv	60 ppmv
	COS/CS_2	Data not available	Ineffective	Data not available	- -	- -	
	R-SH	Data not available	Data not available	Data not available			
Operational, Maintenance Problems, and Reliability		Moderately complex design, simple operation	Corrosion, foaming. Not effective at low pressures. High CO concentrations. Simple design and relatively easy operation.	Must have a minimum of 0.5 to 2.0 percent acid gas. Moderately complex design, simple operation.	Moderately complex design, simple operation.	Solvent cost is 2-3 times the cost of other amines. Foaming, corrosion. Moderately complex design, simple operation.	Inexpensive solvent. Corrosion, foaming
Environmental Considerations:							
System produces an acid gas stream		Yes	Yes	Yes	Yes	Yes	Yes
System produces a solvent blowdown		Yes	Yes	Yes	Yes	Yes	Yes
Preliminary Capital Cost		Approximately 80% of MEA; dependent on the partial pressure of CO_2 and H_2S	\$.81 MM for 20 MM scfd 6% CO_2, 2% H_2S	Approximately 80% of MEA	Comparable to DGA or DEA	Data not available	\$1 MM for 20 MM scfd 6% CO_2, 2% H_2S
Utility Requirements:							
Steam		Yes	Yes	Yes	Yes	Yes	Yes
Electricity		Yes	Yes	Yes	Yes	Yes	Yes
Cooling Water		Yes	Yes	Yes	Yes	Yes	Yes

Source: Ref. 3-1, 3-2, 3-23 through 3-27

TABLE 3.1-7. COMBINATION PHYSICAL/CHEMICAL SOLVENT SYSTEMS AND DIRECT CONVERSION SYSTEM

		Combination Solvent Systems		Direct Conversion
		Amisol	Sulfinol	Stretford
Commercial Application, Purification		Ammonia and methanol	Refinery, natural and synthesis gases, LNG feed stocks, hydrogen	Natural, cokeover and producer gas
Applicability to Gasification		Technically feasible	Technically feasible	Proven
Operating Pressure		Typically 1.38 MPa	atm - 6.90 MPa	Not pressure sensitive
Maximum Control Effectiveness				
High Pressure	H_2S	21 ppmv	<1 ppm	5 ppmv
	COS/CS_2	Total sulfur compounds	<2 ppm	
	R-SH		0.25 grains/100 SCF	Some methymercaptan
Low Pressure	H_2S COS/CS_2 R-SH	Data not available	Data not available	Same as high pressure
Operational, Maintenance Problems, and Reliability		Must have a significant acid gas partial pressure. Operation & design are moderately complex.	Expensive solvent. Operation & design are moderately complex.	A high CO_2 concentration results in lower pH which reduces efficiency. Complex chemistry, moderately complex operation & startup. Acid gas concentration must be less than 15-20 percent.
Environmental Considerations:				
System produces an acid gas stream		Yes	Yes	No
System produces a solvent blowdown		Yes	Yes	Yes
Preliminary Capital Cost		Data not available	Data not available	~$5 MM for 100 MMSCFD of low Btu gas from a coal gasifier gasifying 4% sulfur coal
Utility Requirements:				
Steam		Yes	Yes	Yes
Electricity		Yes	Yes	Yes
Cooling Water		Yes	Yes	Yes

Source: Ref. 3-1, 3-2, 3-23, 3-25, 3-26

of the lack of use in coal gasification facilities, the use of all of the candidate AGR processes in that application is technically feasible (Ref. 3-1, 3-23).

While technically feasible for use in gasification facilities, several of the candidate AGR processes are limited in their application due to the pressure of the gas stream to be treated. The two low-Btu gases listed in Table 3.1-4 are at near atmospheric pressure. These gases could be compressed to moderate pressures (less than 0.7 MPa) without incurring severe economic penalties. However, compression to higher pressures such as 7 MPa (70 atm) would not be economically desirable.

In light of this constraint, several of the candidate AGR processes can be eliminated from consideration for use in a low pressure gasification facility. Included in the list of processes which require high pressures or for which low pressure operation data are not available are all of the physical solvent and combination physical/chemical solvent processes as well as the DEA, DGA, and Benfield processes which use chemical solvents.

The four candidate AGR processes which are effective at near atmospheric or moderate pressures are the MEA, MDEA and DIPA chemical solvent processes and the Stretford processes. The MEA process, operating at slightly above atmospheric pressure ($\sim$0.3 MPa), can achieve levels of 32 ppmv H_2S and 60 ppmv total sulfur (Ref. 3-24, 3-28). The MDEA process can reduce H_2S to 4 ppmv at low pressures (Ref. 3-25), while at 0.5 MPa the DIPA process can obtain H_2S removals to less than 100 ppmv (Ref. 3-25). The Stretford process is not pressure-sensitive, and can reduce H_2S to less than 5 ppmv at low pressures (Ref. 3-1).

In addition to removing H_2S, the three amine processes also remove organic sulfur compounds. Unfortunately, the Stretford process can only remove H_2S (Ref. 3-1). In the amine processes, the organic sulfur absorbed reacts with the solvent to form nonregenerable compounds. In the MEA process this amounts to 15 to 20 percent of the COS removed (Ref. 3-26).

The operation, maintenance, and reliability of the candidate AGR processes are addressed in Tables 3.1-5, 3.1-6, and 3.1-7 with respect to: 1) the corrosive or foaming nature of the solvent, 2) the cost of the solvent, 3) the limiting acid gas concentration, and 4) the process's complexity. As might be expected, these factors vary over a wide range, e.g., from MEA being an inexpensive solvent to the more expensive solvent used in the MDEA process.

Each AGR process generates waste streams that must be disposed of in an environmentally sound manner. For the three amine processes, there are two major waste streams: an acid gas stream and a solvent blowdown stream. The solvent blowdown is generally a very small stream which contains trace pollutants, dissolved NH_3, cyanides, and nonregenerable solvent degradation products. This stream generally requires some type of treatment before disposal. The acid gas stream contains the sulfur compounds removed from the treated gas. The Claus and Stretford processes are sulfur recovery processes commercially used in the petroleum and petrochemical industry to treat the acid gas stream from an AGR process. If a Claus process is used, the Claus tail gas may require additional cleanup to meet environmental regulations. This can be done with one of the commercially available tail gas cleanup processes, such as the Beavon or Scott (Ref. 3-25).

Like the amine processes, the Stretford process produces a solvent blowdown stream which must be treated prior to disposal. However, a significant advantage of the Stretford process is that it not only removes H_2S, but also converts the H_2S removed into elemental sulfur. This feature means that, unlike the amine processes, a separate sulfur recovery process is not required.

Very little general information was available for the preliminary cost impacts of the AGR processes. This is because costs are highly dependent of site specific design parameters such as gas flow rate, operating pressure and temperature, composition of the inlet gas, and desired treated gas specifications. However, where available, general cost information is presented in the tables. The costs reported are all in the same general range. However, since use of the Stretford process does not require installation of a sulfur recovery unit, as do the other AGR processes, it appears that the Stretford may be the lowest cost option. A detailed cost analysis is necessary to confirm this.

Specific data on utility requirements are not available on a common basis for process comparisons. However, all of the processes require cooling water, electricity, and steam.

Based on the information just presented, the Stretford and MEA processes were selected as the "best candidate" AGR processes for use in low-Btu coal gasification facilities. The Stretford process was chosen because of its high H_2S removal efficiency and its ability to convert H_2S into elemental sulfur. The MEA, MDEA, and DIPA processes appear to be comparable with respect to the criteria used. The MEA process was selected because it is more widely used and has sufficient operating and design data available to assess its costs, environmental, and energy impacts.

3.1.3 Summary of "Best Candidate" Low-Btu Gasification Systems

The "best candidate" low-Btu gasification systems selected for detailed analysis are the Wellman-Galusha gasifier with either the Stretford or MEA process used to remove sulfur species from the quenched and cooled low-Btu gas. Table 3.1-8 shows the sulfur content of the two cooled gases that will be treated in the Stretford or MEA processes. Also shown in this table are the maximum residual sulfur levels for each gas in order to meet the target SO_2 emission control levels identified earlier.

The Stretford process can reduce H_2S to below 10 ppmv, but does not remove COS (Ref. 3-1). By examining the COS concentrations of the two gases, the Stretford process is capable of meeting all three emission levels for the low sulfur coal case, but only the moderate control level for the high sulfur coal gas. Thus, the Wellman-Galusha/Stretford system will be examined in detail for both coal feedstocks.

If the cooled low-Btu gas is first compressed to a moderate pressure (0.3 MPa), the MEA process can reduce the total sulfur content of the treated gas to less than 60 ppmv. Since this is below the stringent control level for both the low and high sulfur coal cases, the Wellman-Galusha/MEA system could be considered for all cases.

TABLE 3.1-8. SULFUR CONTENT OF THE TWO BASE CASE LOW-BTU GASES

	High Sulfur Coal Feed	Low Sulfur Coal Feed
Gas Heating Value, MJ/m^3	5.62	5.71
H_2S, ppmv	7000	1100
COS, ppmv	280	54
Maximum Allowable Total Sulfur to Meet Emission Level, ppmv		
- Moderate: 150 ng/J	316	321
- Stringent: 43 ng/J	90	92
- Intermediate: 86 ng/J	181	184

However, the MEA process produces an acid gas stream which must be treated in a sulfur recovery unit. For the high sulfur case, the H_2S concentration of the acid gas stream is estimated to be about 25 vol. percent. This level of H_2S is sufficiently high to allow use of the Claus process for sulfur recovery. Unfortunately, for the low sulfur coal case the H_2S concentration in the acid gas stream is estimated to be only about

6 vol. percent. This level is too low for the acid gas stream to be treated in a Claus process. The major alternative to the Claus is use of the Stretford process which works well on streams with low H_2S concentrations. However, it is not reasonable to examine both the Stretford and the MEA/Stretford as the sulfur removal/recovery systems for the low sulfur coal case. While both systems can meet the stringent SO_2 target control level, preliminary cost estimates indicate that use of the Stretford alone would be much cheaper than use of the MEA process followed by the Stretford process for sulfur recovery.

With respect to NO_x and particulate emissions, all of the low-Btu cases are expected to be able to meet the most stringent target control level. For particulates the target level is 13 ng/J. The estimated grain loading in the quenched low-Btu gases is 0.01 g/m^3 (Ref. 3-19). This is equivalent to around 2 ng/J in the combustion flue gases.

When low-Btu gas is combusted, NO_x emissions from thermal fixation of atmospheric nitrogen are estimated (Ref. 3-20) to be less than or equal to those from natural gas combustion, 50-100 ng/J (Ref 3-29). This emission range is in line with the target NO_x emission control level of 86 ng/J. The amount of NO_x which will be formed from fuel bound nitrogen compounds (NH_3 and HCN) cannot be estimated because the quantities of nitrogen compounds remaining in the cleaned low-Btu are not known. While some will be removed during both the quenching/cooling steps and the acid gas removal step, the amounts removed cannot be accurately estimated.

3.2 SELECTION OF "BEST CANDIDATE" MEDIUM-BTU COAL GASIFICATION SYSTEMS

As for low-Btu gasification systems, it is necessary to select not only "best candidate" medium-Btu gasifiers, but also "best candidate" acid gas removal processes. The candidate gasifiers are compared in Section 3.2.1, while the AGR processes are compared in Section 3.2.2. Section 3.2.3

summarizes the "best candidate" systems, i.e., combinations of gasifiers and AGR processes selected as a result of the comparisons.

3.2.1 Comparison and Selection of a Candidate Medium-Btu Gasifier

The fourteen candidate gasifiers identified in Section 2 and the data which were used to compare them were presented in Table 3.1-2. Information on four of the fourteen candidate gasifiers--Wellman-Galusha, Woodall-Duckham/Gas Integrale, Foster Wheeler/Stoic, and Coalex--indicate that their only application is in the production of low-Btu gas. In addition, information on three other gasifiers--Chapman (Wilputte), Riley Morgan, and Pressurized Wellman-Galusha--indicate that while it is technically feasible to operate them to produce medium-Btu gas, that application has not been commercially demonstrated. Therefore, these seven gasifiers are not considered as candidates for producing medium-Btu gas.

Of the remaining seven candidate medium-Btu gasifiers, four--GFERC Slagging, BGC Slagging Lurgi, BI-GAS, and Texaco--are only in the pilot or demonstration plant stage of development. Moreover, the current development plans for these gasifiers do not indicate that they will be commercially available within the time frame being considered in this study. Thus, only three medium-Btu gasifiers--Lurgi, Winkler, and Koppers-Totzek--can realistically be considered as potential "best candidates" based on their status of development and applicability.

However, the Winkler and Koppers-Totzek gasifiers can be eliminated because they operate at atmospheric pressure. As discussed in Section 2, economic and size considerations indicate that medium-Btu gas will probably only be used if land availability or environmental conditions prohibit the installation of on-site low-Btu gasifiers. Under either of these conditions, off-site generation of medium-Btu gas and pipelining the gas to the boiler is an alternative. But, the economics of transporting gas through a pipeline are improved if the gas is pressurized. And, while it is possible to compress the gas produced by a Winkler or Koppers-Totzek gas, it is less expensive to compress the oxygen feed to a pressurized gasifier than to pressurize the much larger quantity of medium-Btu product gas. Thus, the pressurized Lurgi gasifier is an economic choice over either the atmospheric pressure Winkler or Koppers-Totzek gasifier if the product gas is to be transported.

The estimated compositions of the raw and quenched/cooled medium-Btu gases produced by gasifying the two study coals (see Table 3.1-2) in a Lurgi gasifier are listed in Table 3.2-1. These gas compositions were based on published data for gasifying similar coals (Ref. 3-4, 3-16). Adjustments were made for any differences in coal compositions.

TABLE 3.2-1. ESTIMATED COMPOSITIONS OF MEDIUM-BTU GASES FROM LURGI GASIFIER

	Low Sulfur Coal Feed		High Sulfur Coal Feed	
Composition, Vol. %	Raw Gas	Quenched Gas	Raw Gas	Quenched Gas
CO_2	16.91	29.74	12.91	31.72
CO	12.12	21.32	6.95	17.08
CH_4	7.32	12.88	4.06	9.98
H_2	19.83	34.88	15.51	38.11
H_2O	42.66	0.37	58.94	0.37
N_2	0.30	0.53	0.59	1.45
H_2S	0.16	0.28	0.534	1.31
COS, ppmv	84	150	90	220
NH_3	0.70	tr	0.49	tr
HCN	NA	tr	NA	tr
Tars and Oils, kg/m^3 (@ 43°C, 1 atm)	0.023	0.004	0.012	0.005
HHV, MJ/m^3	6.57	11.6	4.21	10.4

NA - Probably present, but data not available.

tr - Trace.

Entrained particulates and tars/oils in the raw medium-Btu gases produced by the Lurgi gasifier are removed by the quench/cooling steps shown in Figure 3.2-1. The gas is first quenched in a direct contact scrubber and then further cooled in a series of indirect coolers. The gas exiting the last cooler is essentially free of tars, oils, NH_3, HCN and entrained particulates (Ref. 3-4, 3-16), but still contains some light hydrocarbons (C_6 - C_9).

3.2.2 Comparison and Selection of "Best Candidate" Acid Gas Removal Processes

The candidate acid gas removal processes for medium-Btu gasification systems are the same as those for low-Btu systems. Information on the candidate processes was presented in Tables 3.1-5, 3.1-6, and 3.1-7. Comparisons of those processes with respect to their use in a pressurized medium-Btu gasification facility are summarized in the following text. The results of the comparisons are the selection of the Stretford and Rectisol processes as the "best candidate" AGR processes for further study.

As described earlier in Section 3.1.2, all of the candidate processes are commercially available and proven or technically feasible for use in coal gasification facilities. In addition, since the gas from a Lurgi gasifier is at a relatively high pressure (∿2.4 MPa), all fourteen candidate AGR processes can provide adequate removal of sulfur species.

With respect to environmental considerations, all of the candidate processes produce a small solvent blowdown stream which must be treated before being disposed of. All of the processes, except the Stretford, also produce an acid gas stream which must be sent to a separate sulfur recovery process. As discussed previously, the fact that the Stretford process not only removes sulfur species but also recovers the sulfur value, is a significant economic advantage over the other candidate AGR processes which require a separate sulfur recovery process.

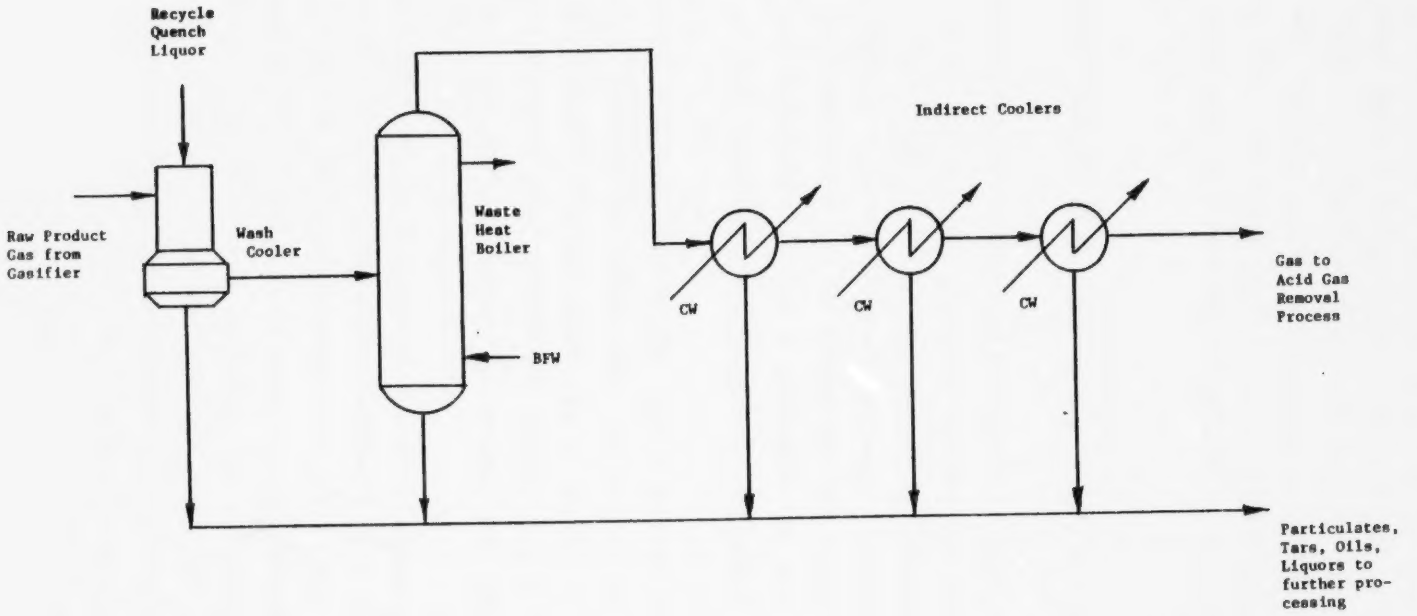

Recycle Quench Liquor
Raw Product Gas from Gasifier
Wash Cooler
Waste Heat Boiler
BFW
Indirect Coolers
CW
CW
CW
Gas to Acid Gas Removal Process
Particulates, Tars, Oils, Liquors to further processing

Figure 3.2-1. Lurgi quench/cooling steps.

Data on capital investment requirements for the candidate AGR processes are limited. The cost information that is available cannot be compared since it is not on a common basis. However, as mentioned above and in Section 3.1.2, the Stretford process appears to be the lowest cost option because it doesn't require a separate sulfur recovery process.

Specific data on utility requirements are not available on a common basis. However, some general comments can be made regarding regeneration steam requirements. The physical solvent processes use depressurization of the rich solvent to release a major portion of the absorbed acid gases. Thus, high inlet CO_2 concentrations and high CO_2 removals do not significantly increase the regeneration steam requirements. On the other hand, the regeneration steam requirements for the chemical solvent processes are directly related to the quantities of acid gases ($CO_2 + H_2S$) removed. This is a significant disadvantage for the chemical solvent processes since the two medium-Btu gases that will be treated have high CO_2 concentrations ($\sim$30 percent). The combination physical/chemical solvent processes also suffer from increased regeneration steam requirements as the CO_2 content of the gas increases. However, since the Stretford process does not remove CO_2, it is not affected by high inlet CO_2 concentrations.

Based on the above discussion, the Stretford and the physical solvent processes are the potential choices for "best candidate" AGR process. As stated at the beginning of this section, the processes finally selected were the Stretford and the Rectisol. The Stretford was selected because it has significant cost advantages. The Rectisol process was selected because 1) it is representative of the chemical solvent processes, 2) it is the only physical solvent process which has had commercial application in gasification plants, and 3) data are available to assess its cost, energy, and environmental impacts.

3.2.3 Summary of "Best Candidate" Medium-Btu Gasification Systems

The "best candidate" medium-Btu gasification systems selected for detailed analysis in Sections 4, 5, and 6 are the Lurgi gasifier with either the Stretford or Rectisol acid gas removal process. For this study, these facilities are assumed to be at an off-site location with pipeline transmission of the medium-Btu gas to the boiler. To avoid pipe corrosion problems (Ref. 3-30), the H_2S content of the fuel gas must be below 10 ppmv. While no information could be found on the maximum permissible level of COS, the Rectisol process normally effects complete removal of COS concurrently with H_2S removal. However, the Stretford process does not remove COS. Based on these sulfur removal levels, the estimated emissions of SO_2 from combusting the four medium-Btu gases would be as follows:

- Low sulfur coal with Rectisol - 2 ng SO_2/J (0.005 lb/10^6 Btu)
- Low sulfur coal with Stretford - 40 ng SO_2/J (0.09 lb/10^6 Btu)
- High sulfur coal with Rectisol - 2 ng SO_2/J (0.005 lb/10^6 Btu)
- High sulfur coal with Stretford - 60 ng SO_2/J (0.14 lb/10^6 Btu)

The first three cases are below the stringent target SO_2 emission control level of 43 ng SO_2/J (0.1 lb/10^6 Btu), while the last case is below the intermediate level of 86 ng SO_2/J (0.2 lb/10^6 Btu).

NO_x emissions from combustion of any of the four medium-Btu gases should be similar to those from natural gas combustion (Ref. 3-20). The EPA emission factor for natural gas (Ref. 3-29) is 50-100 ng/J (0.12-0.23 lb/10^6 Btu), which is comparable to the target NO_x control level of 86 ng/J (0.2 lb/10^6 Btu). Particulate emissions should be negligible since all particulates are removed from the gas during the purification operation.

3.3 SELECTION OF "BEST CANDIDATE" COAL LIQUEFACTION SYSTEMS

As discussed in Section 2, of the coal liquefaction processes being developed, the hydrogenation-type processes have gained in importance in view of the higher conversion efficiencies and lower costs that have been projected by the process developers and various other organizations. The SRC-I, SRC-II, H-Coal, and EDS are hydrogenation-type systems and are the candidate coal liquefaction systems considered in this study.

Solvent Refined Coal (SRC-I) System--The SRC-I system is being developed to convert high-sulfur, high-ash coals to lower-sulfur, low-ash solid fuel. (SRC-II, a variation of the SRC-I system, produces lower-sulfur and low-ash liquid fuels from coal.) Figure 3.3-1 is a schematic of the system. The coal is first pulverized and mixed with a coal-derived solvent in a slurry mix tank. The slurry is combined with hydrogen and is then pumped through a fired preheater and passed into a dissolver. In this unit, the coal is hydrogenated and thereby depolymerized, leading to an overall decrease in product molecular weight and dissolution of the coal. The solvent is also hydrocracked in the dissolver unit yielding lower molecular weight hydrocarbons ranging from light oil to methane. Another reaction occurring in the dissolver is the hydrogenation of the organic sulfur in the coal which produces hydrogen sulfide.

From the dissolver, the mixture passes to a separator where the gases are separated from the slurry of dissolved solids and coal liquids. The raw gas is sent to a hydrogen recovery and gas desulfurization unit. The recovered hydrogen is then recycled and combined with the slurry coming from the slurry mix tank. Hydrocarbon gases are released and recovered and the hydrogen sulfide is converted to elemental sulfur.

The slurry of undissolved solids and the dissolved coal solution is then separated in a filtration unit. In a commercial-scale system, the solids would be sent to a gasifier-converter where they would react with

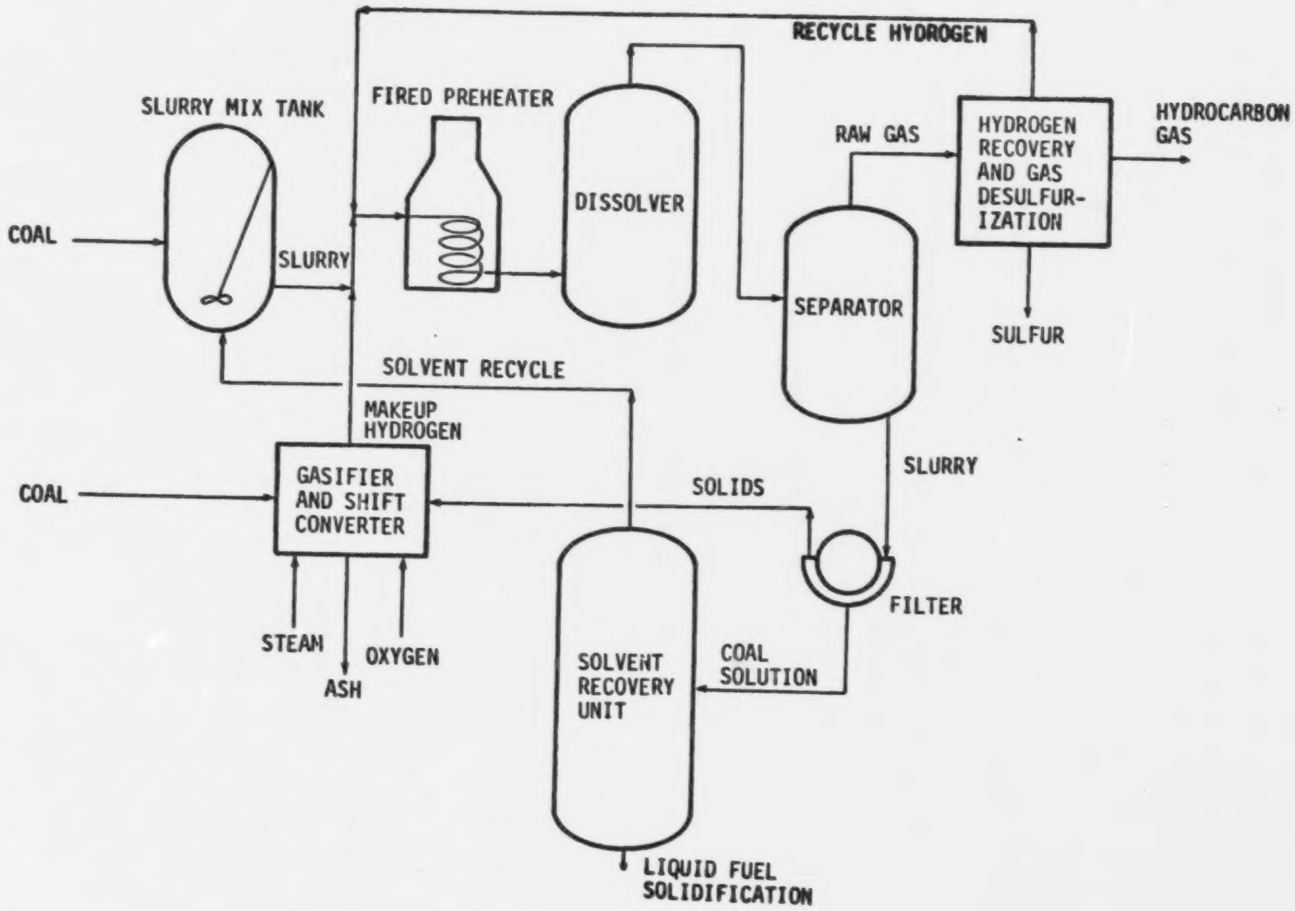

RECYCLE HYDROGEN
SLURRY MIX TANK
FIRED PREHEATER
DISSOLVER
RAW GAS
HYDROGEN RECOVERY AND GAS DESULFUR-IZATION
HYDROCARBON GAS
COAL
SLURRY
SEPARATOR
SULFUR
SOLVENT RECYCLE
MAKEUP HYDROGEN
SLURRY
COAL
GASIFIER AND SHIFT CONVERTER
SOLIDS
FILTER
STEAM
ASH
OXYGEN
SOLVENT RECOVERY UNIT
COAL SOLUTION
LIQUID FUEL SOLIDIFICATION

Figure 3.3-1. Solvent Refined Coal (SRC-I) system.

supplemental coal, steam, and oxygen to produce hydrogen for use in the system. To date, such capability has not been demonstrated on a commercial scale. The coal solution passes to the solvent recovery unit where the solvent is separated from the final liquid product, solvent refined coal. Solvent refined coal (SRC-I) has a solidification point of 175°C to 205°C and a heating value of approximately 37 MJ/kg.

H-Coal System--A schematic of the system is provided in Figure 3.3-2. Coal is crushed to less than 250 μm, dried, slurried with recycled oil, and then pressurized. Compressed hydrogen is added to the slurry and the mixture is preheated and charged continuously to the bottom of the ebullated-bed catalytic reactor. The upward passage of internally recycled reaction mixture keeps the catalyst in a fluidized state. Catalyst activity is maintained by the periodic addition of fresh catalyst and the withdrawal of spend catalyst. The temperature of the ebullated-bed catalytic reactor is controlled by adjusting the preheater outlet temperature.

Vapor product leaving the top of the reactor is cooled to separate the heavier components as a liquid. Light hydrocarbons, ammonia, and hydrogen sulfide are absorbed from the gas stream and sent to a separator and a sulfur recovery unit. The remaining hydrogen-rich gas is recompressed and combined with the input slurry. The liquid from the condenser is fed to an atmospheric distillation unit to separate it into two streams. The overhead light distillate is refined into useful products. The heavy underflow from the still is used as recycled solvent for slurrying feed coal. The liquid-solid product from the reactor, containing unconverted coal, mineral matter, and oil, is fed into a flash separator. The flashed off material is passed to the atmospheric distillation unit which yields light and heavy distillate products. The bottoms product from the flash separator (solids and heavy oil) is further separated with a hydrocyclone, a liquid-solid separator, and by vacuum distillation.

The gas and liquid products, composed of hydrogen gases, hydrogen sulfide, ammonia, light oil distillate, heavy oil distillate, and residual fuel

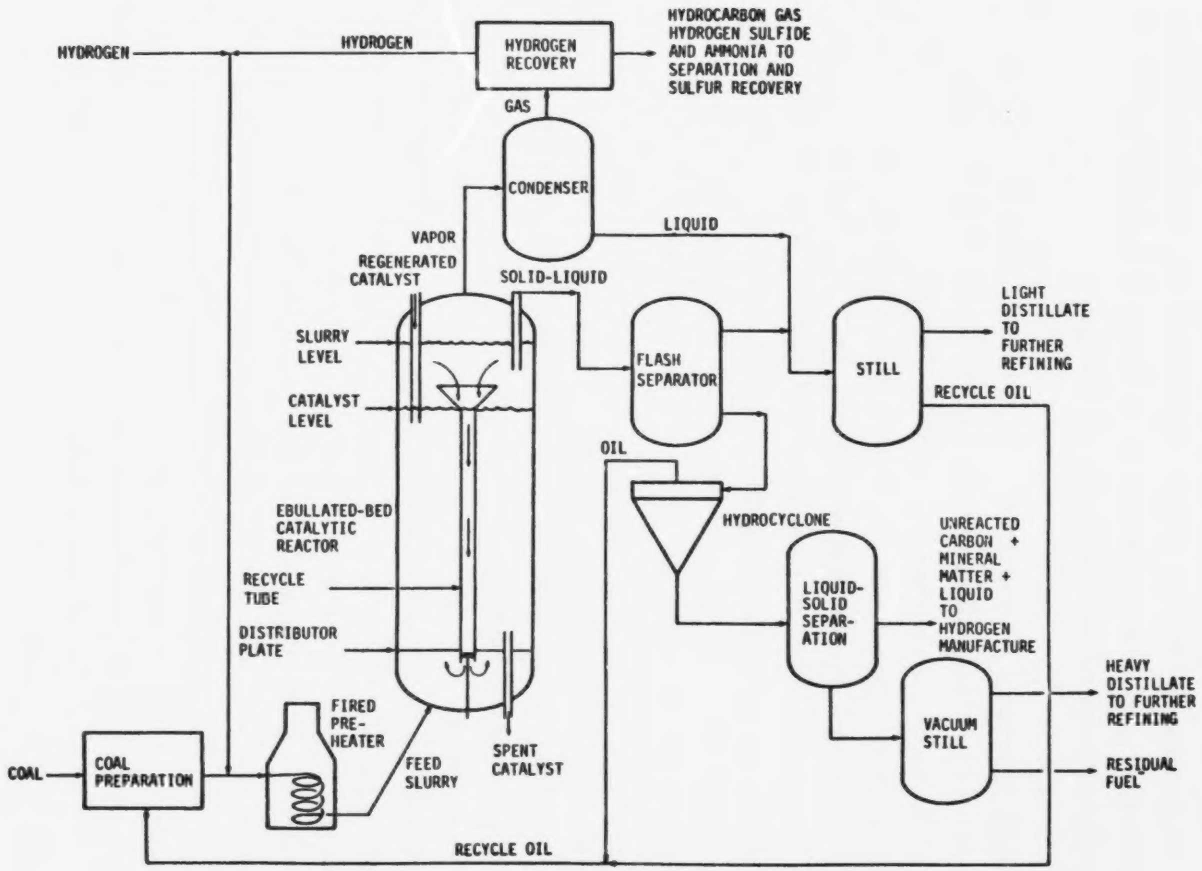

HYDROGEN
HYDROGEN
HYDROGEN RECOVERY
HYDROCARBON GAS HYDROGEN SULFIDE AND AMMONIA TO SEPARATION AND SULFUR RECOVERY
GAS
CONDENSER
VAPOR
LIQUID
REGENERATED CATALYST
SOLID-LIQUID
SLURRY LEVEL
CATALYST LEVEL
FLASH SEPARATOR
STILL
LIGHT DISTILLATE TO FURTHER REFINING
RECYCLE OIL
OIL
EBULLATED-BED CATALYTIC REACTOR
HYDROCYCLONE
UNREACTED CARBON + MINERAL MATTER + LIQUID TO HYDROGEN MANUFACTURE
LIQUID-SOLID SEPAR-ATION
RECYCLE TUBE
DISTRIBUTOR PLATE
HEAVY DISTILLATE TO FURTHER REFINING
VACUUM STILL
FIRED PRE-HEATER
SPENT CATALYST
FEED SLURRY
COAL
COAL PREPARATION
RESIDUAL FUEL
RECYCLE OIL

Figure 3.3-2. H-Coal system.

oil, may be further refined as necessary. A portion of the heavy distillate is recycled as the slurry medium. The stream containing the unreacted carbon and some liquid eventually will be processed in a commercial installation to produce additional hydrogen needed for the process.

Exxon Donor Solvent (EDS) System--A schematic of the system is provided in Figure 3.3-3. The coal is dried and ground to less than 500 μm and mixed with recycled solvent to form a slurry. Hydrogen is normally preheated separately and combined with preheated slurry at the reactor inlet. An alternate mode of operation combines the hydrogen with the slurry before preheating.

After liquefaction, the product is sent to the first stage of separation. Water is removed, hydrogen is recycled, and raw liquid products are sent to the vacuum separation section. The raw liquid product contains unreacted carbon and mineral matter.

The primary vacuum flash tower removes the mineral matter, unreacted coal, and other fractions with boiling points above 540°C (bottoms). The overhead from the first tower is sent to the secondary vacuum tower, to separate heavy oil from the light oil as the overhead. This overhead from the secondary vacuum flash tower is then combined with the lighter liquid stream recovered from the liquefaction reactor separator system and fed to the solvent hydrotreating section, consisting of fixed-bed catalytic reactors. The solvent and naphtha from the solvent hydrotreating section are then separated by fractionation. Most of the solvent is recycled to the slurry vessel. Hydrogen, after undergoing acid gas removal, is also returned to the slurry vessel.

3.3.1 Comparison of the Candidate Coal Liquefaction Systems

The four candidate coal liquefaction systems are compared in Table 3.3-1 with respect to the selection criteria described at the beginning of

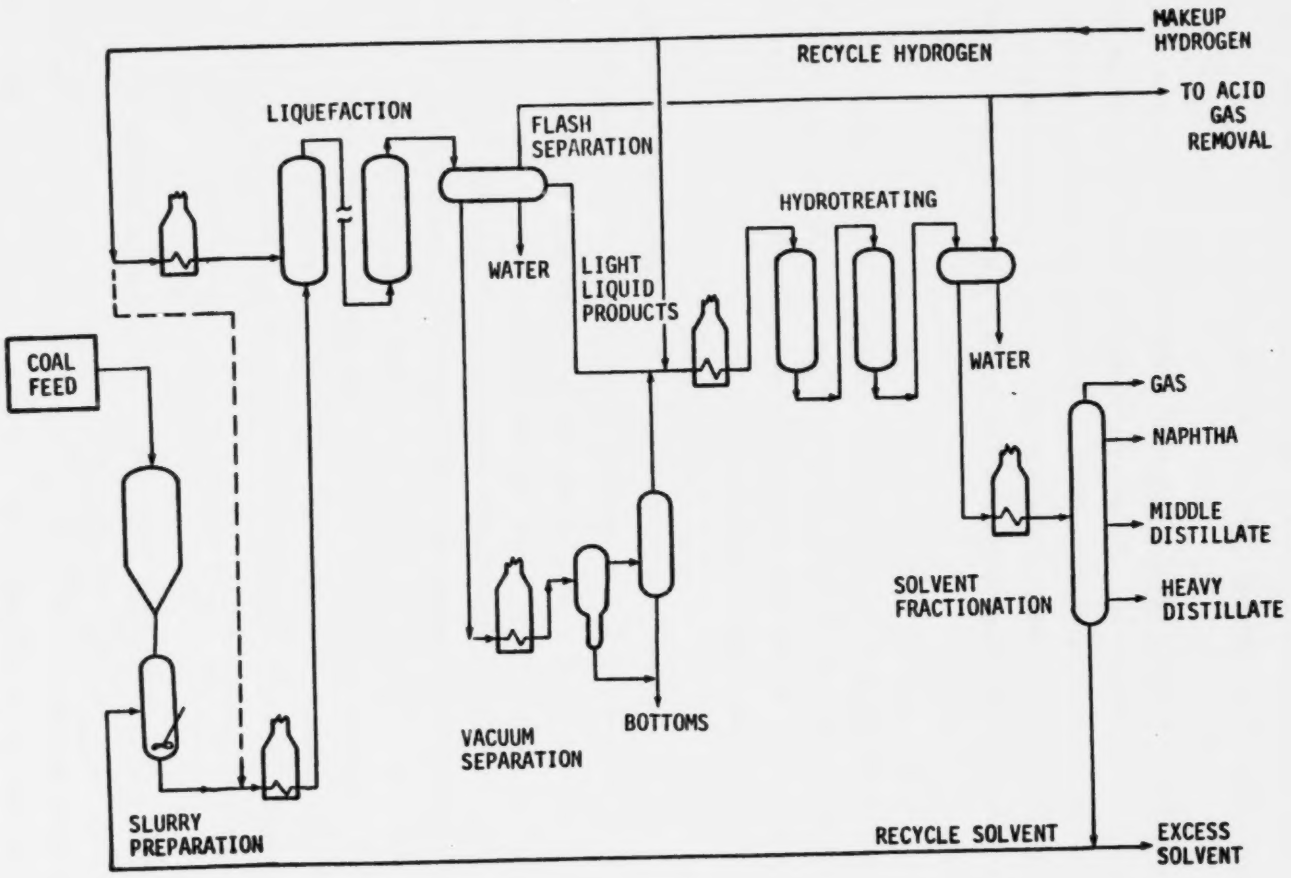

MAKEUP HYDROGEN
RECYCLE HYDROGEN
TO ACID GAS REMOVAL
LIQUEFACTION
FLASH SEPARATION
WATER
LIGHT LIQUID PRODUCTS
HYDROTREATING
WATER
COAL FEED
GAS
NAPHTHA
MIDDLE DISTILLATE
HEAVY DISTILLATE
SOLVENT FRACTIONATION
BOTTOMS
VACUUM SEPARATION
SLURRY PREPARATION
RECYCLE SOLVENT
EXCESS SOLVENT

Figure 3.3-3. Exxon Donor Solvent (EDS) system.

TABLE 3.3-1. CANDIDATE COAL LIQUEFACTION SYSTEMS

Characteristics	SRC-I	SRC-II	H-Coal	Exxon Donor Solvent (EDS)
Status of Development	• 1700 kg/hr pilot plant operating, Ft. Lewis, WA • 210 kg/hr pilot plant, Wilsonville, AL	• 1700 kg/hr pilot plant operating, Ft. Lewis, WA • 210 metric ton/hr demonstration plant planned, West Virginia	• 100 kg/hr PDU operating, Trenton, NJ • 20 metric ton/hr pilot plant under construction, Cattlesburg, KY. Start-up in early 1979	• 0.9 and 2.4 kg/hr bench scale units and 34 kg/hr pilot plant operating, Baytown, TX • 8.6 metric ton/hr under construction, Baytown, TX Operation in 1980
Operational and Maintenance Requirements	• Solids accumulation control in the dissolver on western coals • Solids-liquid separation (filtration) equipment • Slurry pumps	• Solids accumulation control in the dissolver on western coals • Vacuum distillation column • Slurry pumps	• Ebullated bed reactor • Circulating slurry pumps	• Circulating slurry pumps • Flexicoking
Applicability	• Solid fuel produced can only be used in pulverized coal boiler	• Liquid fuel can be used in conventional fuel oil-fired boiler	• Liquid fuel can be used in conventional fuel oil-fired boiler	• Liquid fuel can be used conventional fuel oil-fired boiler
Environmental Impacts*	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes	• Control of sulfur release by sulfur recovery process followed by tail-gas unit • Particulate control from coal preparation • Wastewater treatment to reduce organic and inorganic constituents in waste streams • Disposal of solid wastes
Financial impact (capital cost of the liquefaction plant is prorated to the boiler size)	Capital cost may be as high as 50 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.	Capital cost may be as high as 60-75 percent of the companion utility boiler plant.
Energy Impact**	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%	Thermal efficiency = 60-75%

*See Table 3.3-3, 3.3-4, and 3.3-5.

**Thermal Efficiency is defined as the energy content of all useful products expressed as a percentage of the energy content of the feed coal.

this section. The information contained in this table is summarized in the following text.

The SRC-I, SRC-II, H-Coal, and EDS systems are in an advanced stage of development; however, there are a number of factors such as scale-up difficulties, feedstock flexibility, product utilization, reactor complexity, and cost which are still not well defined on a common basis for these four systems.

Considerable background is available from liquefaction process research programs as to performance of various bench and process development units. The results have shown that yields, hydrogen consumption, and product quality can be duplicated in equipment with a significant range in capacity. Testing of large size pilot plants will further confirm scale-up parameters beyond the smaller size reactor systems that have already been tested. Pilot plants using larger reactor and commercial components and equipment will resolve a number of general issues, in addition to scale-up parameters, that relate to mechanical equipment and process integration. These issues are considered to represent major technical problems to be solved prior to construction of commercial plants.

The various hydrogenation-type coal liquefaction processes use common components and subsystems. Design bases for these components are to be obtained from their use in the larger pilot plants during 1979-1980. A list of subsystems and components required for hydrogenation processes that need further engineering development is given in Table 3.3-2 (Ref. 3-30).

The SRC-II, H-Coal, and EDS systems produce fuel oil-type liquid products that can be used in conventional fuel oil-fired boilers. However, the SRC-I process produces a solid fuel similar to low-sulfur coal except for its low melting point (175°C to 205°C). Because of this, SRC-I is not suitable for use as a fuel in stoker-type boilers. SRC-I is suitable

TABLE 3.3-2. HYDROGENATION SUBSYSTEMS AND COMPONENTS REQUIRING ENGINEERING DEVELOPMENT

Unit Operation	Issues
Slurry mixing and pumping	Erosion and operation of multiple reciprocating pumps
Slurry preheating	Design configuration, gel formation, heat transfer, pressure drop, coking of tubes
Hydrogenation	Contacting of coal, solvent, hydrogen exothermic reactions
Product heat exchange	Fouling of heat exchange surfaces, erosive action of solids
Pressure letdown	Valve life, testing of valves for service
Materials of construction of system units	Long-term durability questionable for service due to corrosion, erosion, embrittlement, and chemical effects
Fractionation	Design of vacuum-still transfer lines, heated pumps, solids carryover, transport of streams with high solids concentration at high temperatures
Hydrogen or fuel gas generation, using pitch containing ash	Feasibility of using gasification or flexicoking for generating hydrogen or fuel gas from vacuum-still bottoms
Operating procedures	Solids withdrawal, startup, emergency shutdown, recovery, and other factors

Source: Ref. 3-30

though for use in a conventional pulverized coal boiler if certain modifications are made (see discussion in Section 2.3.3.1). When commercialized, coal liquefaction systems will be large facilities producing daily the energy equivalent of about 1000 gigajoules of synthetic fuels. This is much more energy than is needed by a single industrial boiler. Therefore, a coal liquefaction plant will only be associated with an industrial boiler as an off-site fuel source transporting a portion of its product to the industrial boiler.

All hydrogenation processes will produce gaseous, liquid and solid waste streams. Tables 3.3-3 through 3.3-5 summarize the sources and characteristics of those streams. The amount and concentration of various compounds present in those wastes will depend on coal characteristics, operating conditions, and hydrogen consumption.

The hydrogen sulfide in the gaseous emissions is removed by an acid gas removal process followed by a sulfur recovery process. Ammonia, formed from coal nitrogen during the hydrogenation reaction step, is removed from the process wastewater by steam stripping.

Phenols in the liquid effluents may be recovered from process wastewater by extraction if their recovery and by-product value is significant. If present in low concentrations they may be treated by biological oxidation treatment. Auxiliary facilities such as raw and boiler water treatment, cooling tower, and coal and product storage generate wastewater effluents and can be treated by conventional methods. Tables 3.3-3, 3.3-4, and 3.3-5 show sources and characteristics of air emission, wastewater streams, and solid waste discharges. Applicability of pollution control methods and equipment will be confirmed by the operation of large-scale pilot and demonstration plants.

For the production of lower sulfur and nitrogen fuels, an additional hydrotreating step may be necessary. More hydrogen sulfide and ammonia is

TABLE 3.3-3. SOURCES AND CHARACTERISTICS OF AIR EMISSIONS

Module	Source	Emission Characteristic
Coal preparation	Grinding, pulverizing and drying	Particulates, hydrocarbon vapors
Hydrogenation	Preheater flue gas	CO_x, NO_x, H_2S, NH_3, hydrocarbons
Pyrolysis and hydrocarbonization	Preheater flue gas	CO_x, NO_x, SO_x, hydrocarbons
Hydrotreating	Preheater flue gas Catalyst removal and replacement	CO_x, NO_x, SO_x, hydrocarbons, particulates, NH_3, H_2S
Catalytic synthesis	Heater flue gas	CO_x, NO_x, SO_x, hydrocarbons
Extraction	None	None
Phase separation	Flash drum vapors	Hydrocarbons, H_2S, NH_3, CO_2
Fractionation	Uncondensed gases from condenser	H_2S, CO_2
Gas cleaning module	CO_2 gas stream	CO_2, H_2S, hydrocarbons
Synthesis gas/hydrogen generation	Synthesis gas Driers flue gas	CO_2, hydrocarbons, SO_x, CO_x NO_x, hydrocarbons
Auxiliary systems and utilities	Cooling tower drift Boiler combustion gases	Biocides, anticorrosive agents, solids, NO_x, CO_x, SO_x, hydrocarbons, fly ash

Source: Ref. 3-31

TABLE 3.3-4. SOURCES AND CHARACTERISTICS OF WASTEWATER STREAMS

Module	Source Description	Wastewater Stream	Constituents
Coal preparation	Coal storage piles, crushing and grinding operations	Storm water runoff	Suspended particles, dissolved solids
Hydrogenation	Cooling and quenching operation	Foul water from quench	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
Pyrolysis and hydrocarbonization	Cooling and quenching operation	Foul water from quench	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
Hydrotreating	Condensing overhead vapors	Condensate	Phenols, ammonia, sulfides
Synthesis gas generation	Cooling and quenching operation	Foul water from quench	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
	Shifting operation	Condensed unreacted water	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
Catalytic synthesis	Condensing overhead vapors	Condensate	Phenols, ammonia, sulfides
Phase separation	Two or three stage pressure reduction	Condensate from overhead condenser	Oils, light hydrocarbons, phenols, ammonia, dissolved sulfides
Fractionation	Cooling overhead vapors	Condensate	Light hydrocarbons, dissolved salts
Gas cleaning	Absorption and regeneration operations	Purge flows	Dissolved sulfides in gas removal solvent
Hydrogen generation	Cooling and quenching operation	Foul water from quench	Phenols, tars, ammonia, thiocyanates, sulfides, and chlorides
	Shifting operation	Condensed unreacted water	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
Supercritical gas extraction	Char quenching operation	Foul water from quench	Phenols, tars, ammonia, thiocyanates, sulfides and chlorides
Auxiliary systems and utilities	Cooling towers and boiler	Blowdown	Dissolved salts
	Plant yard area	Storm water runoff	Suspended particles, dissolved solids, traces of phenols, oil and tars

Source: Ref. 3-31

TABLE 3.3-5. MODULAR SOLID WASTE DISCHARGES

	Solid waste					
Module	Particulate coal	Spent catalyst	Ash/slag mineral matter	Char	Spent (sulfur guard) absorbent (zinc sulfide)	Remarks
Coal preparation	+	0	0	0	0	
Hydrogenation	0	+	0	0	0	Spent catalyst not continuously generated
Pyrolysis/hydrocarbonization	+	0	+	+	0	
Hydrotreating	0	+	0	0	0	Spent catalyst particles in gas or liquid stream
Synthesis gas generation	+	+	+	+	0	Spent catalyst not continuously generated
Catalytic synthesis	0	+	0	0	+	
Extraction (supercritical gas extraction)	0	0	+	0	0	
Phase separations	+	0	+	+	0	Small amounts of unreacted char/oil may be present
Fractionation	+	0	+	+	0	Solids from donor solvent processes only
Gas cleaning (acid gas removal)	0	0	0	0	+	Some systems use sulfur guard absorbents
Hydrogen generation	+	+	+	+	0	
Auxiliary systems utilities	+	0	+	+	0	Particulate product losses during handling, ash and particulates from coal/char burning boilers

+ denotes waste stream is generated in module.
0 denotes waste stream is not generated in module.

Source: Ref. 3-31

thereby generated from the process. While this may be an additional environmental impact at the liquefaction plant site, combustion of the hydrotreated liquid fuels will result in reduced environmental impact at the combustion site.

Cost data for the candidate liquefaction processes are not extensive due to the developing nature of these processes. However, for the SRC-II, EDS, and H-Coal systems capital costs have been estimated at 60-75 percent of the cost of a utility boiler using the liquefied coal products (Ref. 3-30, 3-32). The capital cost of an SRC-I plant is estimated at 50 percent of the associated boiler, but the cost of this boiler may only be 75 percent of the cost of an equivalent conventional coal-fired boiler (Ref. 3-33, 3-34).

Estimates of the thermal efficiency of coal liquefaction plants (defined as product energy content divided by coal feed energy content) are in the 60-75 percent range (Ref. 3-32, 3-35, 3-36). The actual efficiency will depend on factors such as coal characteristics, degree of processing, and environmental requirements.

3.3.2 Selection of "Best Candidate" Coal Liquefaction Systems

Based on the information presented in Section 3.3.1, the SRC-I and EDS systems were selected as the "best candidates" for detailed cost, energy, and environmental analyses. The SRC-I system was chosen because it is the most developed and potentially the most economical of the candidate systems. However, a disadvantage of the SRC-I system which may restrict its use is its inability to produce a very low sulfur fuel from a high sulfur coal feedstock.

The other three liquefaction processes are very similar with respect to status of development, environmental impacts, costs, and energy conversion efficiency. The EDS system was selected because it is representative of the three systems and the results of analyzing the EDS system should also apply to the SRC-II and H-Coal systems.

The sulfur, nitrogen and ash contents and higher heating value of the fuels produced by the SRC-I and EDS systems are shown in Table 3.3-6. Also shown on this table are the estimated emissions of SO_2, and particulates resulting from combustion of those fuels. These emission values were calculated from EPA emission factors (Ref. 3-29) for the combustion of fuel oil and bituminous coal (except for the SRC-I SO_2 value which is based on complete conversion of the fuel sulfur to SO_2). The particulate emissions for SRC-I assume that a particulate control device with at least an 80 percent removal efficiency is applied to the combustion gases. Values for NO_x are shown in parentheses on Table 3.3-6. These are based on EPA emissions factors (Ref. 3-29) which may not be valid for coal derived fuels.

The emission control levels selected for combustion of coal liquefaction fuels are shown below:

	Moderate Level	Intermediate Level	Stringent Level
SO_2, ng/J	520	260	86
NO_x, ng (as NO_2)/J	300	220	86
Particulates, ng/J	13	13	13

In selecting liquefaction processes to meet these emission targets, control of NO_x emissions is not considered. This is because NO_x emissions cannot be estimated "a priori" for coal derived fuels and there is insufficient data available to generalize NO_x emissions from the combustion of these fuels.

The low sulfur coal derived SRC-I and EDS fuels can meet all three target SO_2 and particulate emission levels. The high sulfur coal derived SRC-I fuel can meet all the particulate control levels, but can only meet the moderate SO_2 emission level. The high sulfur coal derived EDS fuel can meet all three target emission levels, but for the stringent level, hydroprocessing is required to reduce the fuel sulfur content to 0.2 percent.

TABLE 3.3-6. FUEL COMPOSITION AND COMBUSTION EMISSIONS FOR LIQUEFACTION FUELS

	SRC-I		EDS		
	Low Sulfur Coal	High Sulfur Coal	Low Sulfur Coal	High Sulfur Coal	
				Raw	Hydroprocessed
Fuel Composition					
% S	0.1	0.8	0.2	0.3	0.2
% N	1.3	1.7	0.2	0.2	0.2
% Ash	0.2	0.3	-	-	-
HHV	36.2 MJ/kg	36.6 MJ/kg	41.4 GJ/m^3	41.4	41.4
Emissions					
SO_2, ng/J	55	440	83	125	83
NO_x, ng NO_3/J	(248)	(248)	(115)	(115)	(115)
Particulates, ng/J	13	13	6	6	6

* Assumes 80% efficient particulate control device on flue gas.

Numbers in parenthesis are merely estimates based on EPA emissions factors (Ref. 3-29). Actual emissions are not known.

References

3-1. Cavanaugh, E. C., W. E. Corbett and G. C. Page. Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology, Volumes I & II, Final Report. EPA 600/7-77-125a and b, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. September 1977.

3-2. Dravo Corp. Handbook of Gasifiers and Gas Treatment Systems. Final Report. Task Assignment No. 4. Report No. FE-1772-11, ERDA Contract No E(49-18)-1772. Pittsburgh, PA. Chemical Plants Div. February 1976.

3-3. Lurgi Mineraloltechnik GmbH. "Lurgi Pressure Gasification Performance Record--Lurgi Express Information". Company Brochure. Frankfurt, Germany. October 1975.

3-4. Woodall-Duckham Ltd. Trials of American Coals in a Lurgi Gasifier at Westfield, Scotland, Final Report. Research and Development Report No. 105, FE 105. Crawley, Sussex, England. November 1974.

3-5. Grant, Andrew J., "Applications of the Woodall-Duckham Two Stage Coal Gasification". Presented at the Third Annual International Conference on Coal Gasification and Liquefaction, School of Engineering. University of Pittsburgh. Pittsburgh, PA. 3-5 August 1976.

3-6. Riley Stoker Corp. "The Riley-Morgan Coal Gasification System". Company Brochure. Worchester, MA. October 1975.

3-7. Rawdon, A. H., R. A. Lisauskas and S. A. Johnson. "Operation of a Commercial Size Riley-Morgan Coal Gasifier". Presented at the American Power Conference. Chicago, Illinois. 19-21 April 1976.

3-8. Lewis, P. S., A. J. Liberatore, and J. P. McGee. Strongly Caking Coal Gasifier in a Stirred-Bed Producer. R.I. 7644. Morgantown, W. Va. Morgantown Energy Research Cntr. 1972.

3-9. Gronhovd, G. H., et al. Design and Initial Operation of a Slagging, Fixed-Bed, Pressure Gasification Pilot Plant. U.S. Bureau of Mines Report of Investigations No. 6085. Washington, D.C., 1962.

3-10. U.S. Energy Research and Development Administration. Quarterly Technical Progress Report, April - June 1976, Grand Forks Energy Research Center. Report No. GFERC/QTR-76/4. Grand Forks, ND. GFERC. September 1976.

3-11. Schora, Frank C., Jr. Fuel Gasification. Advances in Chemistry Series 69, A Symposium Sponsored by the Division of Fuel Chemistry at the 152nd Meeting of the ACS. N.Y. September 1966. Washington, D.C., ACS, 1967.

3-12. Lacey, J.A. "The Gasification of Coal in a Slagging Pressure Gasifier". Amer. Chem. Soc., Div. Fuel Chem., Prepr. 10(4), 151-67. 1966.

3-13. Minnesota, University of. Program Opportunity Notice for the Integration and Evaluation of Low-Btu Coal Gasification Technology in Operational Environments. Volume 1. Technical. Minneapolis, MN. Coal Gasification Duluth Campus Heating Plant. 9 July 1976.

3-14. Kamody, John F., and J. Frank Farnsworth. "Gas from the Koppers-Totzek Process for Steam and Power Generation". Presented at the Industrial Fuel Conference, Purdue Univ., West Lafayette, IN. October 1974.

3-15. Zahradnik, R.L., and R.J. Grace. "Chemistry and Physics of Entrained Coal Gasification". Amer. Chem. Soc., Div. Fuel Chem., Prepr. 18(1), 203-27. 1973.

3-16. El Paso Natural Gas Co. Application of El Paso Natural Gas Co. for a certificate of public convenience and necessity. Docket No. CP73-131. El Paso, TX. 1973.

3-17. Ball, D.A., et al. Environmental Aspects of Retrofitting Two Industries to Low- and Intermediate-Energy Gas from Coal. Final Report. EPA-600/2-76-102. EPA Contract No. 68-02-1843. Battelle Columbus Laboratories. Columbus, OH. April 1976.

3-18. Thomas, W.C., K.N. Trede, and G.C. Page. Environmental Assessment: Source Test and Evaluation Report - Wellman-Galusha (Glen Gery) Low-Btu Gasification. Final Report. EPA Report No. 600/7-79-185, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. August, 1979.

3-19. Page, Gordon C. Environmental Assessment: Source Test and Evaluation Report--Chapman Low-Btu Gasification. Final Report. EPA-600/7-78-202. EPA Contract No. 68-02-2147. Radian Corporation. Austin, TX. October 1978.

3-20. Rawdon, A.H., R.A. Lisauskas and S.A. Johnson. "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames". Presented at the NO_x Control Technology Seminar, sponsored by Electric Power Research Institute. San Francisco, CA. 5-6 February 1976.

3-21. Cheremisinoff, Paul N., and Richard A. Young, eds. "Control of Fine Particulate Air Pollutants: Equipment Update Report". Pollut. Eng. 8(8), 22-29. 1976.

3-22. Calvert, Seymour and Richard Parker. "Collection Mechanisms at High Temperature and Pressure". In Proceedings of the Symposium on Particulate Control in Energy Processes, San Francisco, CA. 11-13 May 1976. Mountain View, CA. Acurex Corp., Aerotherm Div. September 1976. pp. 491-521.

3-23. "NG/SNG Handbook". Hydrocarbon Process. 50(4), 93-122. 1971.

3-24. Perry, Charles R. "Basic Design and Cost Data on MEA Treating Units". In Proc. 1967 Gas Conditioning Conf., University of Oklahoma, Norman, OK.

3-25. Environmental Control Selection Methodology for a Coal Conversion Demonstration Facility. Austin, TX. Radian Corp. February 1978.

3-26. Riesenfeld, F.C., and A.C. Kohl. Gas Purification. Second Edition. Houston, Texas. Gulf Publishing Co. 1974.

3-27. Telephone communication between P.J. Murin (Radian Corp.) and Tom Moore (Jefferson Chemical Co.) concerning performance of DGA Acid Gas Removal Process. July 18, 1978.

3-28. Telephone communication between P.J. Murin (Radian Corp.) and Paul Sigmund (Union Carbide) concerning performance of amine acid gas removal processes. June 5, 1978.

3-29. Environmental Protection Agency. Compilation of Air Pollutant Emission Factors, 2nd ed., with supplements. AP-42. Research Triangle Park, NC, Feb. 1972, April 1973, July 1973, Sept. 1973, July 1974, Jan. 1975, Dec. 1975, Feb. 1976, April 1977.

3-30. Spencer, D.V., S.B. Alpert and R.H. Wolk. Review of Alternative Liquefaction Processes. EPRI: Electric Power Research Institute, Palo Alto, California. 1978.

3-31. Hittman Associates, Inc. Environmental Assessment Data Base for Coal Liquefaction Technology, Vol. I & II, Final Report. EPA-600/7-78-184a & b, EPA Contract No. 68-02-2162. Columbia, MD. September 1978.

3-32. Fant, B.T. Exxon Donor Solvent Coal Liquefaction Commercial Plant Study Design. Prepared for U.S. Energy Research and Development Administration by Exxon Research and Engineering Company: Florham, NJ. January.

3-33. Spencer, Dwain and Oliver Gildersleeve. Market Potential for New Coal Technologies. EPRI Journal, 3(4) 1978.

3-34. Alpert, S.B., et al. Review of Solvent-Refined Coal Technology. Electric Power Research Institute. Palo Alto, CA. March 1978.

3-35. Schmid, B.K. and D.M. Jackson. Recycle SRC Processing for Liquid and Solid Fuels. In: Paper presented at the Fourth Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity. August 2-4, 1977. Pittsburgh, PA.

3-36. "H-Coal Commercial Evaluation Case 1: Fuel Oil Made for 25,000 TPD H-Coal Liquefaction Plant". Fluor Engineers and Constructors, Inc. Prepared for ERDA. March 1976.

SECTION IV

COST ANALYSIS OF SYNTHETIC FUELS FROM COAL SYSTEMS

Synthetic fuels from coal systems considered to be the "best candidates" as emission control techniques for industrial boilers were selected in Section III. Presented in this section are the results of cost analyses of those systems. In developing the cost analyses, boilers with four heat input rates were considered: 8.8, 22, 44, and 58.6 MW_T* (30, 75, 150 and 200 x 10^6 Btu/hr). A limited number of cases using low-Btu gas in a 117 MW_T (400 x 10^6 Btu/hr) boiler were also examined. These boiler sizes correspond to the coal-fired standard boiler sizes being examined in other technology assessment reports being prepared as part of the EPA's Industrial Boiler Program. Therefore, synthetic fuels from coal technologies can be compared to other control techniques for coal-fired industrial boilers. All costs in this report are expressed in terms of mid-1978 dollars.

Synthetic fuels from coal systems are precombustion emission control techniques for industrial boilers, i.e., they supply a "clean" boiler fuel. As such, the approach used for the cost analyses was:

- First develop costs for producing the synthetic fuels,

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*Since Standard International units are being used in this report, megawatts and kilowatts are used to denote both boiler capacity (thermal heat input) and electrical requirements. To minimize potential confusion, MW_T and MW_e will be used to distinguish between megawatts (thermal) and megawatts (electrical), respectively. As an example, a boiler with a heat input capacity of 30 x 10^6 Btu/hr will have an 8.8 MW_T heat input capacity.

- Then incorporate these fuel costs into the cost analyses of the synthetic fuel-fired boilers, and
- Finally compare the synfuel-fired boiler costs to coal-fired boiler costs.

The medium-Btu coal gasification and coal liquefaction systems examined are large capacity, off-site production facilities. They are also complex and expensive installations. As such, it was beyond the scope of this study to perform detailed cost analyses of these systems. Instead, costs were obtained from design studies available in the open literature and adjusted, as appropriate, to reflect differences between the design study bases and the bases used in this study.

For the low-Btu gasification systems, costs for the most expensive equipment items or processing units were obtained from process vendors. Costs for typical and less expensive equipment were estimated from cost correlations such as those presented by Guthrie (Ref. 4-1) and Peters and Timmerhaus (Ref. 4-2).

Capital requirements for the synfuel-fired boilers were based on information developed by PEDCo for the EPA Industrial Boiler Program (Ref. 4-3, 4-4). The basis for the SRC-I-fired boiler and EDS liquids-fired boilers were PEDCo's 58.6 MW_T (200×10^6 Btu/hr) pulverized coal and 44 MW_T (150×10^6 Btu/hr) residual fuel oil boilers respectively. The capital requirement for low and medium Btu gas-fired boilers should be similar to those for natural gas-fired boilers of comparable sizes. Capital requirements for gas-fired boilers were obtained from PEDCo (Ref. 4-4).

A general discussion of the various components which are included in the cost analyses are presented in Section 4.1. Also presented are the common key assumptions and cost bases used. The results of the cost analyses, assumptions and bases specific to each synfuel technology, and a discussion of significant findings are presented in Sections 4.2 through 4.4 Section 4.2 addresses low-Btu gasification, while medium-Btu gasification is addressed in Section 4.3 and coal liquefaction in Section 4.4. Section 4.5 provides an overall summary of the synfuel economics.

4.1 CONTRIBUTORS TO CONTROL COSTS AND COST BASES

Capital requirements and operating costs for the gasification and liquefaction systems are presented in this section. Additionally, the capital requirements are annualized so that a synfuel cost (in \$/GJ or \$/10^6 Btu) can be calculated. Finally the annualized costs of firing the synfuels in new industrial boilers are developed so that comparisons to the costs for direct firing of coal can be made. The bases for the economic evaluation are presented below.

4.1.1 Capital Requirement

The capital requirement for each synfuel system is presented with the following components identified:

- Installed equipment costs (for each major processing area)

- Indirect costs
  - Engineering costs
  - Construction and field expenses
  - Contractor's fee
  - Startup and performance tests

- Allowance for funds used during construction

- Contingency

- Total turnkey costs (the sum of the items above)

- Working Capital

- Land
- Total capital requirement (the sum of all items listed above)

The installed equipment costs include the cost of all equipment (process and auxiliary) as well as the cost of installation. Indirect capital costs are those costs not attributable to any specific piece of equipment and include the items listed above. Generally, the indirect cost items are taken as a percentage of the installed equipment costs. Specific percentages used in this study are indicated on the cost summary sheets in Sections 4.2 through 4.4 and in Appendices A, B, and C.

If the construction period is long (>1 year), interest during construction on the capital spent in the early stages of construction can become significant. Since a large synfuel plant will require a 3-4 year construction period, allowances must be made for the capital tied up during construction.

The synfuel facilities described in this report have not been built and integrated with gas-fired boilers on a commercial scale. Therefore, economics cannot be compared with or based on an existing facility. A large contingency is, therefore, justified for these systems. A 20% contingency was applied to the uncontrolled coal-fired boiler capital investments by PEDCo (Ref. 4-3, 4-4). It is recommended that a 30% contingency factor be applied to the synthetic fuels system capital investment.

The total turnkey cost (TTC) is the sum of the installed equipment cost, indirect costs, contingencies, and interest during construction. The TTC is also the depreciable investment. Working capital (taken as 25% of total direct operating costs - to be defined in Section 4.1.2) and land are capital requirements but cannot be depreciated for tax purposes.

4.1.2 Operating Costs

The operating costs associated with the synfuels plants are broken out as follows:

- Direct Operating Costs
  - Utilities
    - Steam
    - Electricity
    - Cooling water
    - Process water
  - Raw Materials
    - Coal
    - Water
    - Catalysts and chemicals
  - Operating Labor
  - Maintenance
  - Supervision
  - By-product credits
    - Sulfur
    - Hydrocarbons
    - Ammonia
  - Waste disposal

- Indirect Operating Costs
  - Plant overhead
  - Payroll overhead

Unit prices for the various operating cost components are shown in Table 4.1-1.

4.1.3 Annualized Costs

Annualized costs are the sum of direct operating costs, indirect operating costs, and capital-related costs. Included in the capital-related costs in this study are:

- Local taxes
- Insurance

TABLE 4.1-1. VALUES USED FOR ANNUAL COST ITEMS

Item	Cost[a]	
Operating Labor, \$/man-hour	12.02	
Supervision, \$/man-hour	15.63	
Maintenance Labor, \$/man-hour	14.63	
Steam, \$/GJ (\$/10^6 Btu)	5.00	(5.28)
Electricity, mills/kW-hr	25.8	
Purchased Water, \$/m^3 (\$/10^3 gal)	0.032	(0.12)
Process Water, \$/m^3 (\$/10^3 gal)	0.04	(0.15)
Cooling Water, \$/m^3 (\$/10^3 gal)	0.05	(0.18)
Coal, \$/metric ton (\$/short ton)		
High Sulfur Eastern	18.72[b]	(17.00)
Low Sulfur Western	8.88[b]	(8.06)
Plant and Payroll Overhead[c]		
Waste disposal[d], \$/metric ton (\$/short ton)		
Low Btu gasification facilities	44	(40)
Medium Btu gasification facilities	11	(10)
Coal liquefaction facilities	11	(10)
By-product credits	See section 4.3	

[a] Mid-1978 dollars.

[b] These coal costs are representative of minemouth costs for the Eastern and Western coals. The effect of higher coal costs (considering assumed coal transportation costs) will be explored in a sensitivity analysis for each synfuel technology.

[c] For this study, plant overhead was taken as 26 percent of the annual labor and maintenance costs. Payroll overhead was taken as 30 percent of the annual operating labor costs.

[d] Large gasification and liquefaction facilities will take advantage of large, on-site waste disposal systems and hence have lower solids disposal costs.

Source: Ref. 4-5

- General and administrative (G&A) costs
- Capital recovery
- Interest of working capital

Local taxes, insurance, and G&A are taken as four percent of total turnkey costs (TTC).

The capital recovery factor (CRF) used in this study is based on the equal payment annuity formula:

$$\frac{i\ (1 + i)^n}{(1 + i)^n - 1}$$

where i = annual interest rate

n = economic life of equipment, in years.

For the EPA's Industrial Boiler Program, all contractors are using a 10 percent interest rate which will yield the following CRF's:

Economic Life, Years	CRF
20	0.1175
25	0.1102
30	0.1061
45	0.1014

The CRF is applied only to the depreciable investment (Total Turnkey Costs). Interest on working capital is calculated using a 10% interest rate.

4.2 LOW-BTU COAL GASIFICATION

Annualized costs for industrial steam generation facilities using low-Btu Wellman-Galusha gasifiers and low-Btu gas-fired boilers are summarized in Table 4.2-1. These gasification/steam generation facilities are alternatives to direct coal-fired boilers, and are effective in reducing flue gas

TABLE 4.2-1. COST ANALYSIS OF "BEST CANDIDATE" LOW-BTU COAL GASIFICATION SYSTEMS FOR INDUSTRIAL BOILERS[a]

Boiler heat input, MW_T (10^6 Btu/hr)	Coal feed to gasifier	Sulfur control technique and level of control	SO_2 control efficiency, %	Annualized Low-Btu Gas-Fired Boiler Costs		% increase in costs over uncontrolled coal-fired boiler costs[b]
				10^3/yr	per unit of heat input, \$/kW_T (\$/10^6 Btu/hr)	
8.8 (30)	Low-sulfur	Stretford/Stringent	94.2[c]	2100	239 (70,000)	115
8.8 (30)	High-sulfur	Stretford/Moderate	94.2[d]	2320	264 (77,200)	144
8.8 (30)	High-sulfur	MEA/All Levels	94.2-98.2[e]	2540	289 (84,700)	168
22 (75)	Low-sulfur	Stretford/Stringent	94.2[c]	2890	131 (38,500)	54
22 (75)	High-sulfur	Stretford/Moderate	94.2[d]	3380	154 (45,100)	83
22 (75)	High-sulfur	MEA/All Levels	94.2-98.2[e]	3870	176 (51,600)	110
44 (150)	Low-sulfur	Stretford/Stringent	94.2[c]	4450	101 (29,700)	42
44 (150)	High-sulfur	Stretford/Moderate	94.2[d]	5340	121 (35,600)	73
44 (150)	High-sulfur	MEA/All Levels	94.2-98.2[e]	6050	138 (40,300)	97
58.6 (200)	Low-sulfur	Stretford/Stringent	94.2[c]	5320	91 (26,600)	21
58.6 (200)	High-sulfur	Stretford/Moderate	94.2[d]	6530	111 (32,700)	54
58.6 (200)	High-sulfur	MEA/All Levels	94.2-98.2[e]	7350	125 (36,800)	74
117.2 (400)	Low-sulfur	Stretford/Stringent	94.2[c]	9340	80 (23,400)	18
117.2 (400)	High-sulfur	Stretford/Moderate	94.2[d]	11,810	101 (29,700)	53

[a] Wellman-Galusha gasifier; mid-1978 dollars; 60% annual operating factor

[b] Low-Btu gas-fired costs were compared to uncontrolled coal-fired boiler costs for systems using the same coal feedstock and having identical boiler heat inputs.

[c] SO_2 emission level is 30 ng/J (0.07 lb/10^6 Btu)

[d] SO_2 emission level is 140 ng/J (0.32 lb/10^6 Btu)

[e] SO_2 emission levels are 140-43 ng/J (0.32-0.10 lb/10^6 Btu)

All systems have estimated NO_x and particulate emissions as follows:

NO_x--50-100 ng/J (0.12-0.23 lb/10^6 Btu)
particulates--<4 ng/J (<0.01 lb/10^6 Btu)

emissions of SO_2, NO_x, and particulates compared to the coal-fired systems. Because the removal of sulfur species is the most important and expensive gas purification operation, the costs summary and subsequent figures and tables report costs as a function of the degree of control of sulfur species. However, use of low-Btu gas also results in nominally 100 percent control of particulate emissions. This is because essentially all particulates are removed from the low-Btu gas (grain loading of 0.01 g/m^3 which is equivalent to 2 ng/J in the combustion flue gases) during the operations needed for the removal of sulfur species. NO_x emissions are similar to NO_x emissions from the combustion of natural gas--50-100 ng NO_x (as NO_2)/J. Details on the emissions resulting from low-Btu gasification/steam generation facilities are presented in Section 6.1.

Section 4.2 is organized as follows:

4.2.1 Costs Summary and Analysis
4.2.2 Cost Bases
4.2.3 Cost Sensitivity Analysis
4.2.4 Confidence Intervals for Annualized Costs

4.2.1 Costs Summary and Analysis

As described in Chapter 3.1, two processes have been selected for the removal of sulfur species from low-Btu fuel gas. These processes are the Stretford process and the monoethanolamine (MEA) process. The Stretford process features the direct conversion of H_2S to elemental sulfur; organic sulfur compounds are not removed (Ref. 4-6). The MEA process features the absorption of both H_2S and organic sulfur compounds, with subsequent treatment of an acid gas product stream requisite. In this report, subsequent treatment of the acid gas occurs in Claus and SCOT units.

The annualized costs for low-Btu gasification/steam generation facilities have been compared to the annualized costs for coal-fired boilers (without emission controls) that have been prepared by PEDCo (Ref. 4-3). These cost comparisons were made for steam boilers having identical heat input (not steam output). The percentage increase in annualized costs for the low-Btu systems compared to coal-fired systems is summarized in the final column of Table 4.2-1. The largest percentage increase occurs in the smaller systems. The annualized uncontrolled coal-fired boiler costs used for the comparisons are listed in Table 4.2-2.

A breakdown of the cost components for all of the low-Btu gasification/ steam generation facilities studied is provided in Appendix A. Tables 4.2-3 through 4.2-5 are examples of one set of tables contained in Appendix A. Table 4.2-3 summarizes the capital requirements of the gasification systems, while annualized operating costs for the gasification systems (60 percent operating factor) are summarized in Table 4.2-4, along with the estimated product gas cost. Table 4.2-5 presents the total annualized operating costs for the low-Btu gas-fired boilers burning the low-Btu gas with economics developed in Tables 4.2-3 and 4.2-4.

In order to facilitate the evaluation of the control costs associated with low-Btu gasification, several figures were prepared.

- Figure 4.2-1 presents the annualized boiler costs, in $\$/kW_T$, as a function of SO_2 emission level. As shown in the figure, the costs for the MEA system do not vary discernibly with varying SO_2 control efficiency.

- Figure 4.2-2 presents the incremental costs of emissions control (by low-Btu gasification) as a function of boiler size. Because of greater economies of scale in the low-Btu gasification/steam generation systems (versus the direct coal-fired steam generation systems), large boiler systems have markedly lower unit control costs.

- Figure 4.2-3 grafically illustrates the cost effectiveness (incremental annualized costs per kg SO_2 controlled) of low-Btu gasification/steam generation systems in controlling SO_2 emissions.

TABLE 4.2-2. ANNUALIZED COST FOR UNCONTROLLED COAL-FIRED INDUSTRIAL BOILERS

Industrial Boiler			Annualized Costs	
Input Heat Rate, MW_T (10^6 Btu/hr)	Type	Coal Feed	$\$10^3$/yr	$\$/kW_T$ ($\$/10^6$ Btu/hr)
8.8 (30)	Underfeed Stoker	High Sulfur Eastern	952	108 (31,700)
8.8 (30)	Underfeed Stoker	Low Sulfur Western	977	111 (32,600)
22 (75)	Chain-Grate-Stoker	High Sulfur Eastern	1,851	84 (24,700)
22 (75)	Chain-Grate-Stoker	Low Sulfur Western	1,866	85 (24,900)
44 (150)	Spreader Stoker	High Sulfur Eastern	3,075	70 (20,500)
44 (150)	Spreader Stoker	Low Sulfur Western	3,121	71 (20,800)
58.6 (200)	Pulverized Coal	High Sulfur Eastern	4,248	72 (21,200)
58.6 (200)	Pulverized Coal	Low Sulfur Western	4,369	75 (21,800)
117.2 (400)	Pulverized Coal	High Sulfur Eastern	7,784	66 (19,500)
117.2 (400)	Pulverized Coal	Low Sulfur Western	7,930	68 (19,800)

Mid-1978 dollars.

Assumptions: 1) 60% annual operating factor
2) 10% interest rate and the following equipment lifetimes were used to annualize capital requirements
- 8.8 MW_T - 30 years
- 22, 44, 58.6, and 117.2 MW_T - 45 years

Source: Ref. 4-3

TABLE 4.2-3. CAPITAL INVESTMENT REQUIREMENTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
INSTALLED EQUIPMENT COSTS					
Coal Receiving and Storage	370	390	440	600	790
Gasification System	500	1,000	2,000	2,500	5,000
Gas Purification System					
Blower/Compressor	60	70	100	120	240
Quench/Cooling Towers	50	80	120	140	280
Pumps (for cooling liquor)	30	40	70	90	170
Electrostatic Precipitator	120	220	290	370	740
Separator/Evaporator	120	190	240	250	500
H_2S Removal Unit	1,330	1,380	1,800	1,970	2,300
Sulfur Recovery Unit	-	-	-	-	-
Building and Ductwork Not Costed in Above Equipment Costs (6% of EC)	160	220	320	390	640
Total Installed Equipment (EC)	2,740	3,590	5,380	6,430	10,660
INDIRECT INSTALLATION COSTS					
Engineering (10% of EC)	270	360	540	640	1,070
Construction and Field Expense (10% of EC)	270	360	540	640	1,070
Construction Fees (10% of EC)	270	360	540	640	1,070
Startup (2% of EC)	50	70	110	130	210
Total Indirect Costs (IC)	860	1,150	1,730	2,050	3,420
CONTINGENCIES (30% of EC + IC)	1,080	1,420	2,130	2,540	4,220
TOTAL TURNKEY COSTS (TTC)	4,680	6,160	9,240	11,020	18,300
WORKING CAPITAL (25% of Total Direct Operating Costs)	170	270	440	560	1,020
TOTAL CAPITAL INVESTMENT	4,850	6,430	9,680	11,580	19,320

10^3 mid-1978 dollars.

TABLE 4.2-4. ANNUALIZED COSTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
DIRECT OPERATING COSTS					
Operating Labor (@ $12.02/hr)	210	260	320	370	580
Supervision (@ $15.63/hr)	70	70	70	70	140
Maintenance (6% of Total Turnkey Costs x 0.8)	220	300	440	530	880
Replacement Parts (included in maintenance)	-	-	-	-	-
Electricity (@ 25.8 mills/kW-hr)	20	40	80	110	210
Steam (@ $5.00/GJ)	10	30	60	80	160
Cooling Water (@ $0.05/$m^3$)	10	30	60	80	170
Process Water (@ $0.04/$m^3$)	-	-	-	-	-
Coal (@ $8.88/ton)	90	220	440	600	1190
Chemicals	20	50	110	150	300
Sulfur, Ash Disposal (@ $44.00/ton)	30	90	170	230	460
Total Direct Operating Costs	680	1090	1750	2220	4090
OVERHEAD COSTS					
Payroll (30% of operating labor)	60	80	100	110	170
Plant (26% of labor, parts and maintenance)	130	160	220	250	420
Total Overhead Costs	190	240	320	360	590
CAPITAL CHARGES					
G&A, Local Taxes and Insurance (4% of Total Turnkey Costs)	190	250	370	440	730
Capital Recovery (11.75% of Total Turnkey Costs)	550	720	1090	1290	2150
Interest on Working Capital (@ 10%)	20	30	40	60	100
Total Capital Charges	760	1000	1500	1790	2980
TOTAL ANNUALIZED COSTS	1630	2330	3570	4370	7660
Average Gas Cost, $/GJ	9.80	5.60	4.30	3.95	3.45

10^3 Mid-1978 dollars

TABLE 4.2-5. ANNUALIZED COSTS FOR LOW-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
CAPITAL INVESTMENT, 10^3 \$					
Total Turnkey Costs (TTC)	640	1110	1740	2070	4150
Land	<10	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs, excluding fuel costs)	70	70	120	120	200
Total Capital Requirement	710	1180	1860	2190	4350
OPERATING COSTS, 10^3 \$/yr					
Direct Costs					
Operating Labor (@ \$12.02/hr)	110	110	210	210	320
Supervision (@ \$15.63/hr)	70	70	70	70	70
Maintenance Labor (@ \$14.63/hr)	30	30	60	60	130
Replacement Parts	30	40	60	70	140
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50	100
Process Water (@ \$0.04/m^3)	<10	<10	<10	<10	10
Fuel	1630	2330	3570	4370	7660
Chemicals	<10	<10	10	10	10
Total Direct Operating Costs	1900	2620	4030	4840	8440
Overhead					
Payroll (30% of operating labor)	30	30	60	60	100
Plant (26% of labor and parts)	60	70	100	110	170
Total Overhead Costs	90	100	160	170	270
Capital Charges					
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80	170
Capital Recovery Factor (10.61% of TTC)	70	120	180	220	440
Interest on Working Capital (@ 10%)	10	10	10	10	20
Total Capital Charges	110	170	260	310	630
TOTAL ANNUALIZED COSTS, 10^3 \$/yr	2100	2890	4450	5320	9340

Mid-1978 dollars

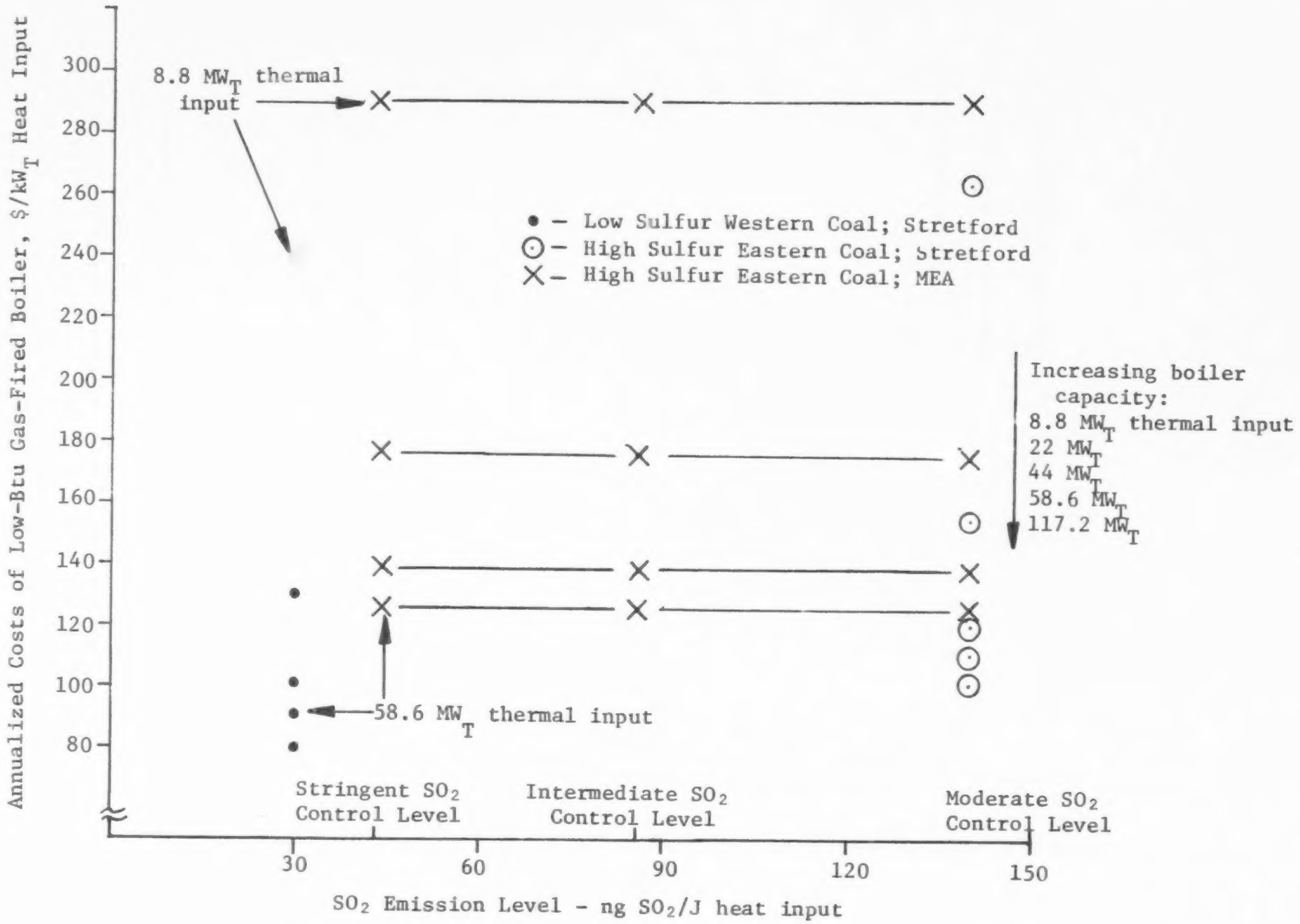


Figure 4.2-1. Annualized costs of low-Btu gasification/steam generation system versus boiler SO$_2$ emission level.

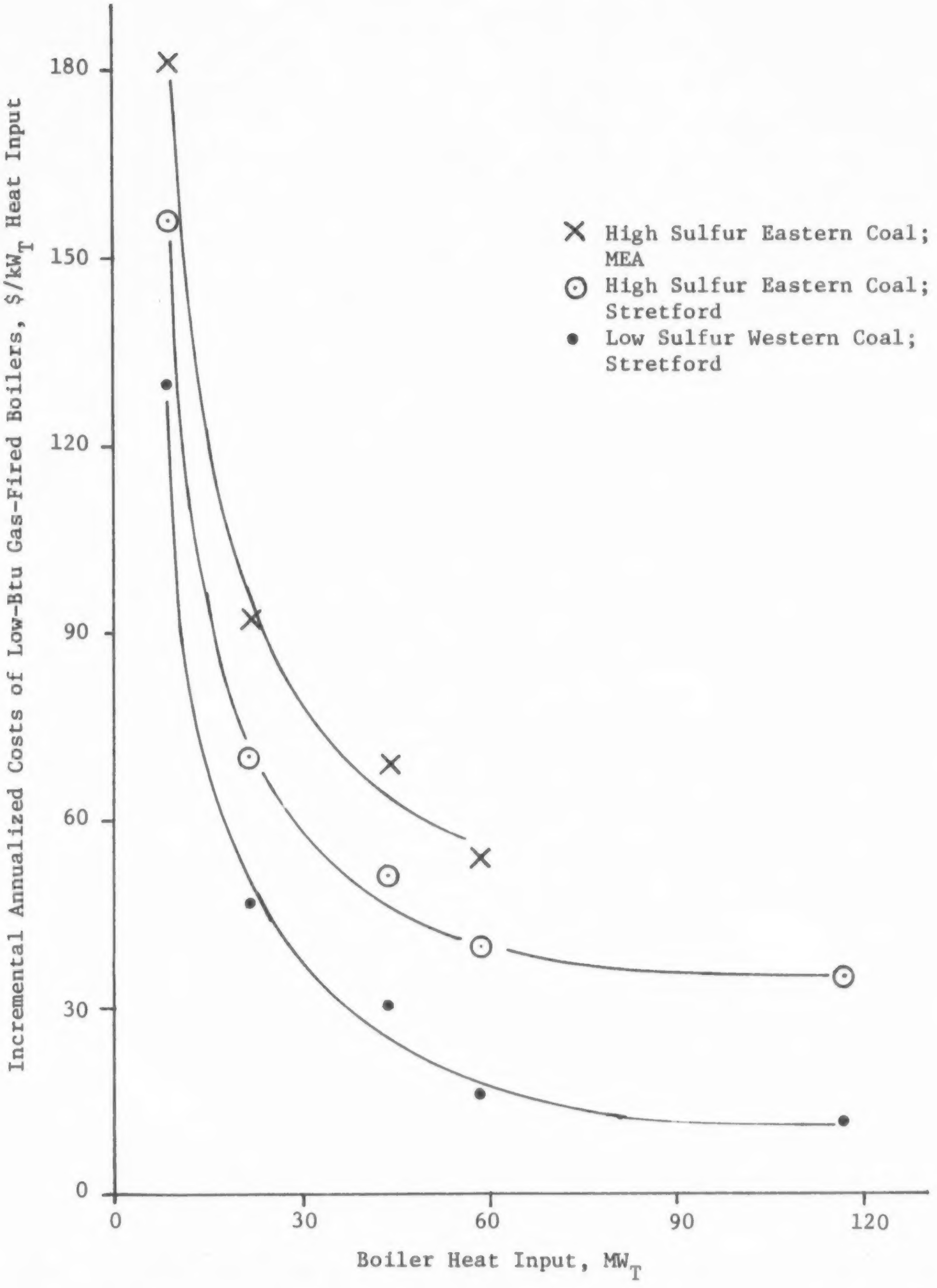
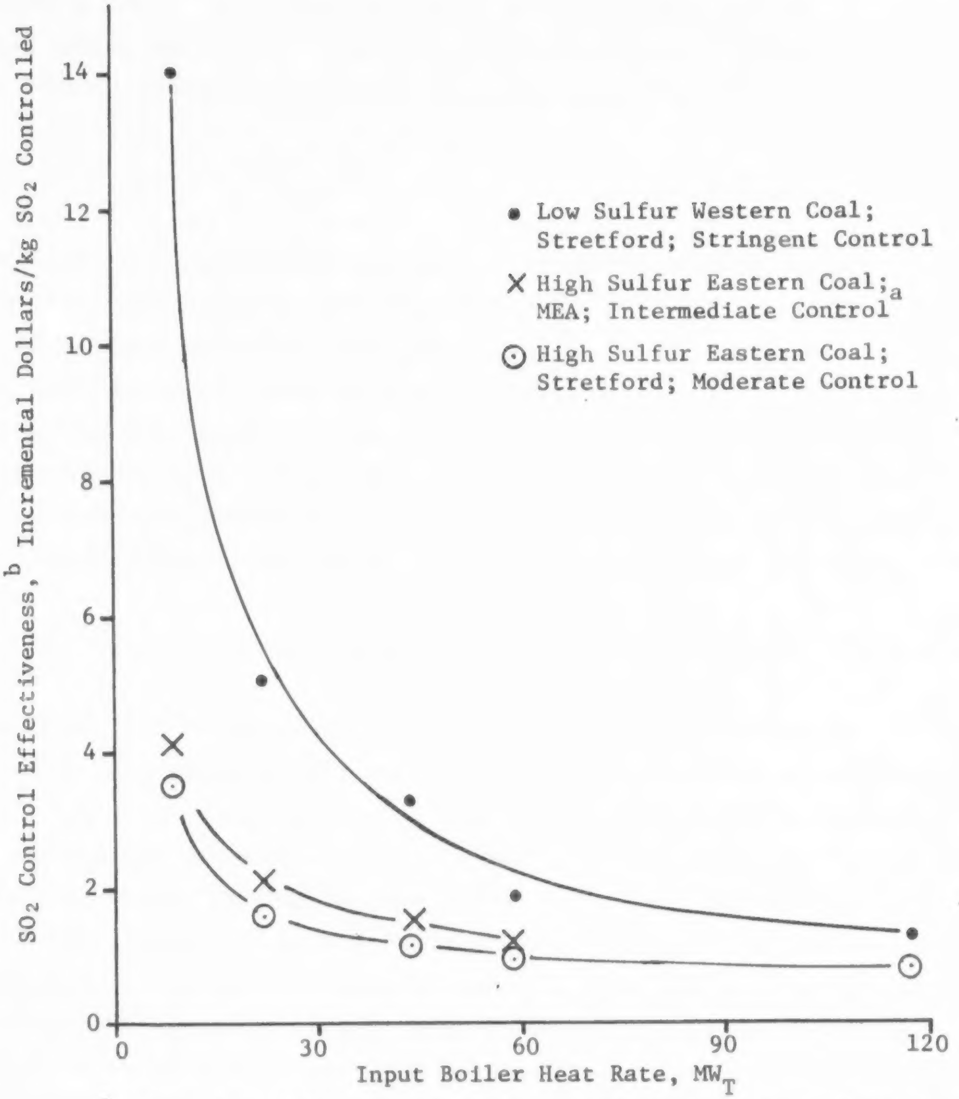


Figure 4.2-2. Incremental annualized costs of low-Btu gas-fired boilers versus boiler input heat rate.



[a] Values for Moderate and Stringent SO_2 Control have less than a 5% variation from the intermediate control value.

[b] Assumes that total incremental control costs are allocated to SO_2 removal.

Figure 4.2-3. Cost effectiveness of low-Btu gasification/steam generation systems in controlling SO_2 emissions.

The following are concluded: 1) SO_2 removal from high sulfur eastern coal using the Stretford unit is more cost effective than using an MEA unit, and 2) SO_2 removal from high sulfur eastern coal is more cost effective than from low sulfur western coal.

4.2.2 Cost Bases

Costs for the most expensive equipment in the gasification/steam generation facility, e.g., the Stretford unit and the gasifiers, were obtained from process vendors. Gas-fired boiler capital costs were provided by PEDCo (Ref. 4-4). Costs for typical and less expensive equipment, e.g., ductwork, were estimated from cost correlations, such as those presented by Guthrie (Ref. 4-1) and Peters and Timmerhaus (Ref. 4-2). Some costs were estimated from those presented in PEDCo's report for uncontrolled coal- and gas-fired boilers (Ref. 4-3). All costs have been indexed to mid-1978 dollars.

4.2.2.1 Cost Basis for Low-Btu Gasification Systems

Costs for the Stretford process were provided by J.F. Pritchard and Co. (Ref. 4-7) and quoted as turnkey costs. (The Stretford, in these instances, produces a sulfur cake product that must be disposed of.) J.F. Pritchard and Co. estimated the costs from cost curves based on their experience, and reported the costs to be no better than preliminary estimates (Ref. 4-7). For the smaller applications, no accurate cost data exist, and costs were extrapolated from costs for larger systems. A ten percent engineering cost and 30 percent contingency were added to J.F. Pritchard's costs to account for overall contractor engineering costs and uncertainties in the cost estimates. Operating costs for the Stretford process include cooling water, chemicals, and electricity. For a system producing 8.8 MW_T of low-Btu gas using high-sulfur coal, these cost items amount to 0.29 m^3/min (77 gpm) of cooling water, $210/d for chemicals, and 63 kW_e (Ref. 4-8, 4-9).

Cost for the MEA absorption unit are based on data reported by Perry Gas Company (Ref. 4-10, 4-11). Because systems of equal capacity but different removal requirements have about the same sorbent flow rate, the same skid-mounted package unit can be used for each level of control. Operating costs include electricity and steam, as described in Chapter 5.1. These amount to 0.9 kW_e electric power and 0.44 MW_T (1.5 x 10^6 Btu/hr) steam for a system with a gas output of 8.8 MW_T and an SO_2 control efficiency of 94.2 percent.

Costs for the Claus and SCOT units were obtained from data of Ralph M. Parsons and Co. and Shell Oil (Ref. 4-12 through 4-16). As for the cost estimates for the Stretford process, costs for the small units were extrapolated off cost curves. Both the Claus and SCOT units for the two smallest boiler systems are smaller than those currently operating, and actually resemble pilot-scale units. The accuracy of the cost data are thus somewhat questionable. Although the acid gases from the MEA units differ slightly (due to the differing levels of control), their impact on Claus/SCOT units is negligible. Utilities for Claus/SCOT units are summarized in Chapter 5.1. For a gasification/steam generation system of 8.8 MW_T thermal input, the utilities will amount to 7 kW_e electric power, 0.022 m^3/min (6 gpm) cooling water, 17 kW_T (5.8 x 10^4 Btu/hr) fuel gas, and steam credit of 0.08 kg/s (630 lb/hr).

Turnkey costs for Wellman-Galusha gasifiers, cyclones, and gasification auxiliaries were obtained from McDowell-Wellman Co. (Ref. 4-17). As was done for the Stretford unit costs, a ten percent engineering cost and 30 percent contingency were added to McDowell-Wellman Co's. cost estimates. The smallest facility in this study (30 MW_T) requires one 3-meter (10-foot) diameter gasifier; the largest (117.2 MW_T) needs ten gasifiers. Power requirements for gasifier jacket water circulation and the gasifier inlet air blower amount to about 38 kW_e for a gasifier/boiler using low-sulfur coal (8.8 MW_T input to boiler).

Installed costs for ductwork and buildings and services were estimated to be 6 percent of the installed equipment investment, based on factors reported in Peters and Timmerhaus (Ref. 4-2).

Costs for the coal receiving and storage system were estimated from costs reported by PEDCo and EBECO (Ref. 4-4, 4-18).

Costs for the tar/oil electrostatic precipitators were obtained from Research Cottrell, Inc. (Ref. 4-19). For a gasifier/boiler facility using low-sulfur coal (8.8 MW_T heat input to boiler), electric power requirements amount to 6 kW_e (Ref. 4-19).

The gasification systems described in this report feature either a centrifugal blower or compressor downstream of the gas cooling equipment and upsteam of the acid gas removal equipment. Those systems using the Stretford process for H_2S removal produce a gas that enters the Stretford process at a pressure of about 110 kPa (16 psia). Those systems using the MEA process for the removal of sulfur compounds produce a gas that enters the MEA process at about 275 kPa (40 psia). Thus, systems using the Stretford process use a blower while systems using the MEA process use a compressor. The estimation of compression power requirements (in kW_e) is summarized in Chapter 5.1. Costs of the blower were estimated from Peters and Timmerhaus (Ref. 4-2), while costs for the compressor were estimated from Guthrie (Ref. 4-1).

Tars and heavy oils are separated from the cooling liquor in a tar/oil/water separator. The separator provides for gravity settling of tars and oils, and provides for their periodic removal and possible utilization. The separator costs are relatively minor and are estimated from Ref. 4-20.

Excess process condensate is evaporated in multiple-effect evaporators to reduce wastewater discharge. After calculating the area of the evaporator, costs were estimated by using factors from Guthrie (Ref. 4-1). The area was calculated from

$$\text{Area} = \frac{\text{steam usage}}{(\text{overall heat transfer coefficient})\ (\text{log mean temperature difference})}$$

As an example, with a heat transfer coefficient of 300 $Btu/hr \cdot ft^2 \cdot {}^\circ F$, a temperature difference of 106°F, and a steam rate of 0.12 MW_T (0.42×10^6 Btu/h), the area of each of three evaporators is calculated to be approximately 13 ft^2, with an installed cost per evaporator of $26,000.

Costs for the cooling liquor circulation pumps were also estimated from charts in Guthrie (Ref. 4-1). The charts use a capacity factor to determine the equipment cost of pumps; the factor is expressed as gpm x psi. As an example, a pump with a circulation of 230 gpm and a pressure drop of 100 psi has a capacity factor of 23,000. Total direct and installation costs for pump and driver amount to $13,000. For a dual set of pumps and drivers (one operating and one spare) the total installed equipment costs are $26,000. Electricity requirements for the pump are estimated in Section 5.1.

Costs for the direct contact cooling towers were based on calculations from Ref. 4-21 and 4-22, and costing charts from Guthrie (Ref. 4-1). The area of the cooling towers was estimated from the maximum allowable vapor velocity to avoid entrainment. The height of the towers was established through heat transfer analysis described in Ref. 4-21 and 4-22. Most tar and oil particles (>1 μm) are removed in the spent cooling liquor.

Unit costs for the operating cost items identified above were presented in Table 4.1-1. The cost of maintenance labor and supplies was calculated as 6 percent of the total turnkey costs and corrected by a factor of 0.80 to account for the reduced operating factor of 60 percent (Ref. 4-2). Operating labor, supervision, and maintenance labor for the various system are presented in Table 4.2-6.

No credit or debit has been taken for tars and oils produced from the gasifiers. This assumes that the value of these by-product hydrocarbons will offset the cost of locating a buyer and transporting the liquids to that buyer. Additionally, the by-product sulfur is not of high quality and is disposed of as a solid waste.

TABLE 4.2-6. LABOR REQUIREMENTS FOR LOW-BTU GASIFICATION/ STEAM GENERATION

Boiler Heat Input, MW_T	8.8	22	44	58.6	117.2
Gasification System					
Operating Labor[a]	2	2½	3	3½	5½
Supervision[a]	½	½	½	½	1
Maintenance Labor[b]	-	-	-	-	-
Gas-fired Boiler					
Operating Labor[a]	1	1	2	2	3
Supervision[a]	½	½	½	½	½
Maintenance[a]	¼	¼	½	½	1

[a] Values shown are number of persons required per 8-hr shift. Total labor costs are based on 3 shifts per day.

[b] Maintenance labor costs for gasification systems are not broken out separately from the annual maintenance materials costs.

4.2.2.2 Cost Basis for Low-Btu Gas-Fired Boilers--

Capital requirements for natural gas-fired boilers were provided by PEDCo (Ref. 4-4). Boilers firing low-Btu gas should be very similar to those firing natural gas, and as a result similar costs are expected. For the purposes of this study, the costs of a low-Btu gas-fired boiler was assumed to be identical to that of a natural gas-fired boiler. All boiler costs are for single, package units except for the largest size, 117 MW_T (400×10^6 Btu/hr) thermal input. For this size, two 58.6 MW_T (200×10^6 Btu/hr) package units were used to avoid the higher costs associated with a field erected unit which would otherwise be required (Ref. 4-4). Labor requirements for the low-Btu gas-fired boilers are presented in Table 4.2-6. Other operating costs were based on information provided by PEDCo (Ref. 4-3, 4-4).

Annualized capital requirements for the low-Btu gas-fired boilers were calculated using the equal payment annuity formula. The 0.1061 capital recovery factor used was based on a 10 percent interest rate and 30-year equipment lifetime.

4.2.3 Cost Sensitivity Analysis

In this section, the major cost components of the low-Btu gas-fired boiler annualized costs are identified. Specifically the effect of coal costs on annualized costs is evaluated since the coal costs studied are typical of minemouth coal costs and some degree of coal transportation and handling is likely for most industrial applications.

A complete cost component breakdown of the annualized cost for low-Btu gas-fired boilers is presented in Table 4.2-7.

Since the coal-to-gas conversion in the gasifier is less than unity, an increase in coal cost will affect the annualized costs of a low-Btu gas-fired boiler to a greater extent than the annualized costs of a coal-fired boiler. Examples of the extent of this impact are presented in Table 4.2-8. For

TABLE 4.2-7. SENSITIVITY ANALYSIS--LOW-BTU GAS-FIRED BOILER ANNUALIZED COSTS

Coal	Low Sulfur Western		High Sulfur Eastern	
Boiler Input Heat Rate, MW_T	8.8	58.6	8.8	58.6
	Percent of Annualized Costs			
Labor and Maintenance	11.4	7.7	10.3	6.3
Utilities	1.4	1.1	1.3	0.9
Overhead	4.3	3.2	3.9	2.6
Capital Charges	5.2	5.8	4.7	4.7
Fuel (Low-Btu gas)				
Labor & Maintenance	23.8	18.2	22.4	15.6
Coal	4.3	11.3	6.9	16.5
Other Operating Costs	4.3	12.2	6.9	16.8
Overhead	9.0	6.8	8.6	5.8
Capital Charges	36.2	33.6	34.9	30.6
Total Fuel	77.6	82.1	79.7	85.3
Annualized Costs, 10^3\$/year	2100	5320	2320	6530

Basis: 1) A Stretford is the acid gas removal system for all cases shown.

2) The first four items in the table above - labor and maintenance, utilities, overhead, capital charges - refer to the gas-fired boiler system. The fifth item - fuel - is low-Btu gas from the Wellman-Galusha System. Percentages under fuel are the various cost components of the gasification system.

TABLE 4.2-8. EFFECT OF COAL TRANSPORTATION COSTS ON ANNUALIZED LOW-BTU GAS-FIRED BOILER COSTS[a]

Input Heat Rate to Boiler, MW_T (10^6 Btu/hr)	Coal Cost = Base Case Values[b]			Coal Transportation Costs Added[c]		
	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Low-Btu Gas-fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Low-Btu Gas-Fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr
Low Sulfur Western Coal						
8.8 (30)	980	2100	1120	1140	2330	1190
58.6 (200)	4370	5320	950	5480	6830	1350
High Sulfur Eastern Coal						
8.8 (30)	950	2320	1370	1040	2430	1390
58.6 (200)	4250	6530	2280	4800	7260	2460

[a] All gasification facilities use the Stretford process

[b] Base Case Values: Low Sulfur Western (LSW) - \$0.40/GJ; High Sulfur Eastern (HSE) - \$0.68/GJ

[c] Transportations Costs: LSW - \$1.00/GJ; HSE - \$0.50/GJ

illustrative purposes coal transportation costs are taken as $1.00/GJ and $0.50/GJ for low sulfur western and high sulfur eastern coal, respectively.

4.2.4 Confidence Interval for Low-Btu Gasification/Boiler Annualized Costs

In assessing the accuracy or confidence interval for the annualized low-Btu gas-fired boiler costs presented in this section, two major items were considered:

- capital investment requirements, and
- operating and maintenance requirements (including labor, maintenance, utilities, raw materials, and waste disposal)

Additional factors impact the accuracy of the annualized costs, but those factors were essentially assumptions or bases imposed on the economic evaluation. Included in the assumptions and bases were:

- method of annualizing the capital investment requirements,
- method of calculating overhead costs, and
- unit costs for operating and maintenance (O&M) requirements

Annualized capital related charges and overhead costs can vary widely depending upon individual company operations. In order to estimate these charges in a manner consistent with the other control technologies being examined in the EPA's Industry Boiler Program, the EPA stipulated the estimating technique to be used. Also, unit costs for the O&M requirements were provided by the EPA in order to maintain consistency in the Industry Boiler Program. Since the above three items served as the bases for the economic evaluation, they were not considered in assessing the accuracy of the annualized cost estimates.

Assessing the accuracy of the capital investment and O&M requirements is complicated by the fact that no low-Btu gasification system, integrated with quenching/cooling and sulfur species removal units, has ever been built.

Thus, there is no "yardstick" to compare the results of this study to. However, the process vendors who supplied the major equipment item cost estimates (e.g., for the gasification system and Stretford process) indicated that the estimates had an accuracy of about ±30%. A conservative approach was taken in estimating the total turnkey costs of the gasification system by including a 30% project contingency. As a result, the total capital investment requirement estimates are believed to be accurate within ±30%.

The operating and maintenance requirements for the gasification systems (e.g., labor and utilities) were estimated based on vendor quotes, information in the literature, and engineering calculations. As such, the accuracy of the O&M requirements are believed to be within ±30%.

Thus, for the assumptions and bases used in this study (e.g., units costs for coal, labor, utilities, and waste disposal), the annualized costs for the low-Btu gas-fired boilers are estimated to be accurate to within ±30%.

4.3 MEDIUM-BTU COAL GASIFICATION

As described in Section 3.2, two processes have been selected in this study for the removal of sulfur species from medium-Btu gas produced by a Lurgi gasifier. These are the Stretford process and the Rectisol process. Also, use of two coals--a high sulfur eastern and a low sulfur western (see Section 3 for coal analyses)--are being examined. A large centralized gasification plant with a medium-Btu gas capacity of 3060 MW_T (250×10^9 Btu/SD) was the only plant size investigated. Therefore, the medium-Btu gas cost (in $/GJ) is constant for all of the medium-Btu gas-fired boilers examined in this study. Thus, four coal gasification system scenarios are considered and costs for four medium-Btu gas products were developed. The estimated annualized operating costs for industrial boilers using these medium-Btu gases as fuel are summarized in Table 4.3-1.

As discussed in Section 6.2 on environmental impacts, the levels of particulate and NO_x emissions from the medium-Btu gas-fired boilers are

TABLE 4.3-1. COST ANALYSIS OF "BEST CANDIDATE" MEDIUM-BTU COAL GASIFICATION SYSTEMS FOR INDUSTRIAL BOILERS*

Boiler Heat Input, MW_T (10^6 Btu/hr)	Coal Feed to Gasifier	Sulfur Control Technique**	SO_2 Control Efficiency, %	Annualized Medium-Btu Gas-fired Boiler Costs			% Increase in Annualized Costs Over Uncontrolled Coal-Fired Boiler Costs
				$\$10^3$/yr	per unit heat input $\$/kW_T$	($\$/10^6$ Btu/hr)	
8.8 (30)	Low-sulfur	Stretford	93	1210	138	(40,400)	24
8.8 (30)	Low-sulfur	Rectisol	99+	1320	150	(43,900)	35
8.8 (30)	High-sulfur	Stretford	98	1490	169	(49,500)	57
8.8 (30)	High-sulfur	Rectisol	99+	1650	188	(55,100)	73
22 (75)	Low-sulfur	Stretford	93	2390	109	(31,900)	28
22 (75)	Low-sulfur	Rectisol	99+	2690	122	(35,700)	44
22 (75)	High-sulfur	Stretford	98	3120	142	(41,600)	69
22 (75)	High-sulfur	Rectisol	99+	3500	159	(46,600)	89
44 (150)	Low-sulfur	Stretford	93	4550	103	(30,200)	46
44 (150)	Low-sulfur	Rectisol	99+	5150	117	(34,300)	65
44 (150)	High-sulfur	Stretford	98	6010	137	(40,100)	95
44 (150)	High-sulfur	Rectisol	99+	6770	154	(45,100)	120
58.6(200)	Low-sulfur	Stretford	93	5840	100	(29,300)	34
58.6(200)	Low-sulfur	Rectisol	99+	6630	113	(33,100)	52
58.6(200)	High-sulfur	Stretford	98	7770	133	(39,000)	83
58.6(200)	High-sulfur	Rectisol	99+	8790	150	(43,900)	107

*Lurgi gasifier. 60% annual operating factor for the boiler.

**All cases have estimated NO_x and particulate emissions of 50-100 ng NO_x/J (0.12-0.23 lb/10^6 Btu) and 4 ng particulates/J (0.01 lb/10^6 Btu).

SO_2 emission levels (from the industrial boiler only) are as shown below:

Low sulfur western coal with Stretford - 37 ng SO_2/J (0.09 lb/10^6 Btu)
High sulfur eastern coal with Stretford - 60 ng SO_2/J (0.14 lb/10^6 Btu)
All Rectisol cases - 2 ng SO_2/J (0.004 lb/10^6 Btu)

estimated to be similar to those for natural gas combustion--50-100 ng NO_x/J (0.12-0.23 lb/10^6 Btu) and 4 ng particulates/J (0.01 lb/10^6 Btu). This is esentially complete control of particulates and 60-85 percent control of NO_x emissions from what would be expected from direct-firing of coal. Sulfur dioxide emissions are also significantly reduced by firing medium-Btu gas instead of coal, with reductions of 93 to 99+ percent being achieved. A more detailed discussion of the environmental impact of using medium-Btu gasification as an industrial boiler emission control technique is presented in Section 6.2.

This section is organized as follows:

4.3.1 Costs Summary and Analysis
4.3.2 Cost Bases
4.3.3 Cost Sensitivity Analysis

4.3.1 Costs Summary and Analysis

In Table 4.3-1, the annualized costs for medium-Btu gasification/steam generation facilities are compared to the annualized costs for uncontrolled coal-fired boilers (See Table 4.2-2 for descriptive information on the coal-fired boiler systems). Refer to Section 4.3.2 for the economic bases for this evaluation. As indicated by the "percentage increase" data in Table 4.3-1, the medium-Btu gas-fired boilers are most nearly competitive with uncontrolled coal-fired boilers at the small boiler capacities. This occurs since the medium-Btu gas cost does not vary as a function of boiler capacity. However, as will be shown in Figure 4.3-2, the incremental annualized costs of the medium-Btu gas-fired boilers (versus coal-fired boilers) per kW_T of design input heat to the boiler do not have a significant dependence on boiler input heat rate.

In order to facilitate the evaluation of the emissions control costs associated with medium-Btu gasification, Figures 4.3-1, 4.3-2, and 4.3-3 were prepared.

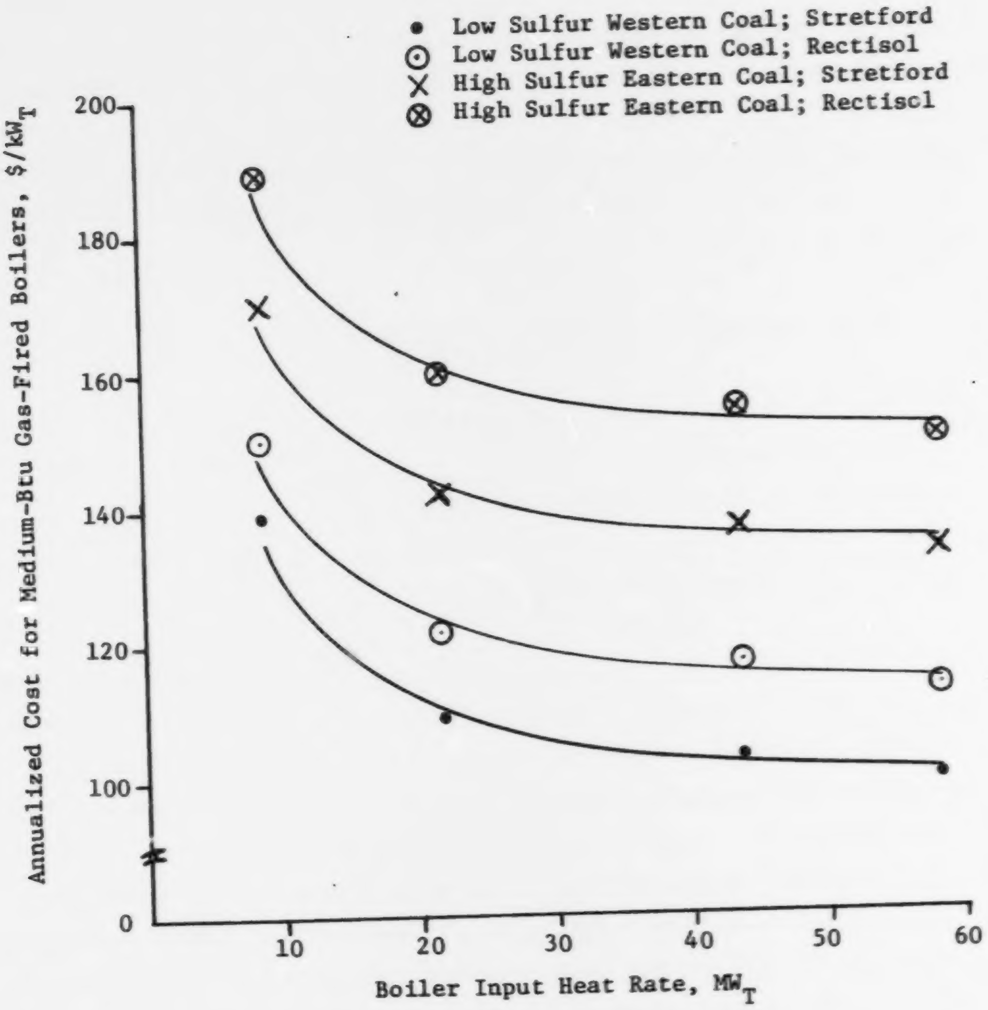


Figure 4.3-1. Annualized costs of medium-Btu gas-fired industrial boilers.

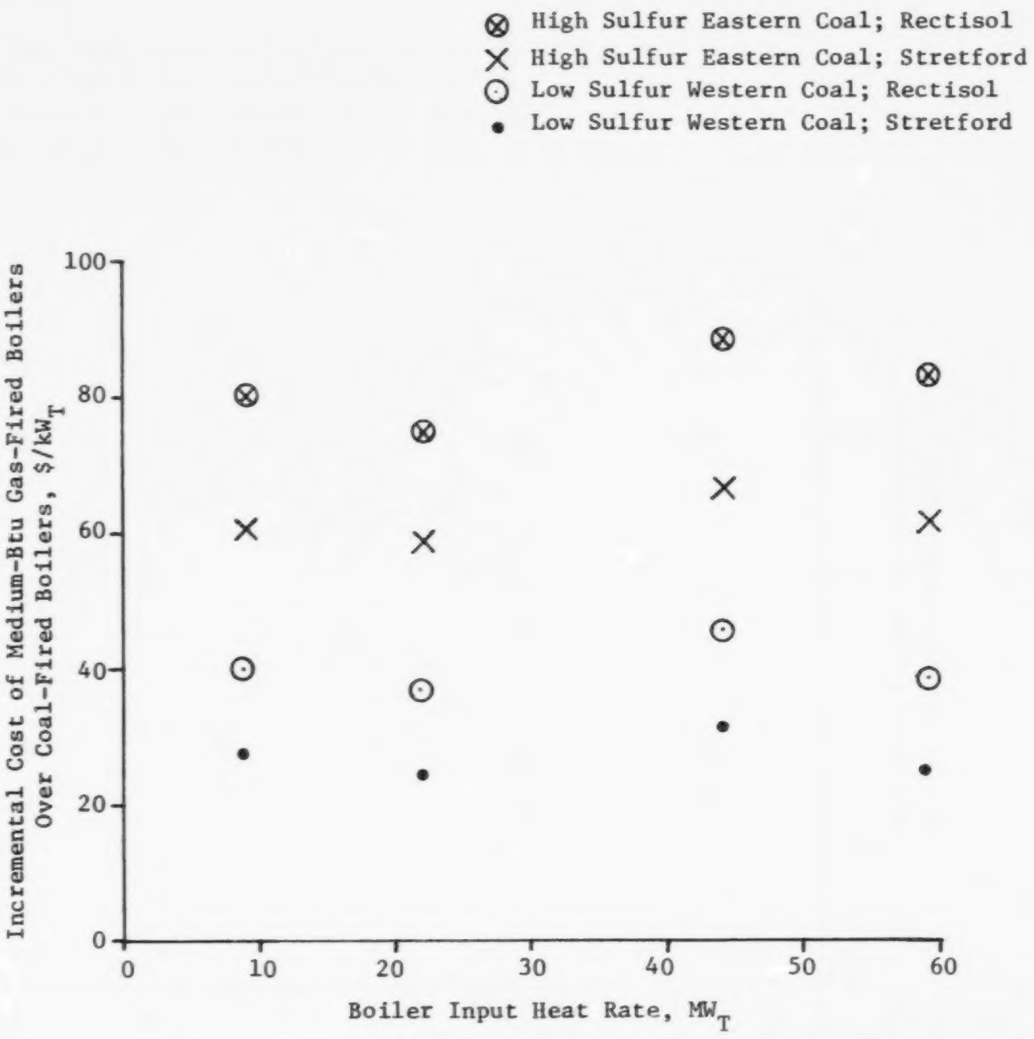


Figure 4.3-2. Incremental cost of medium-Btu gas-fired industrial boilers

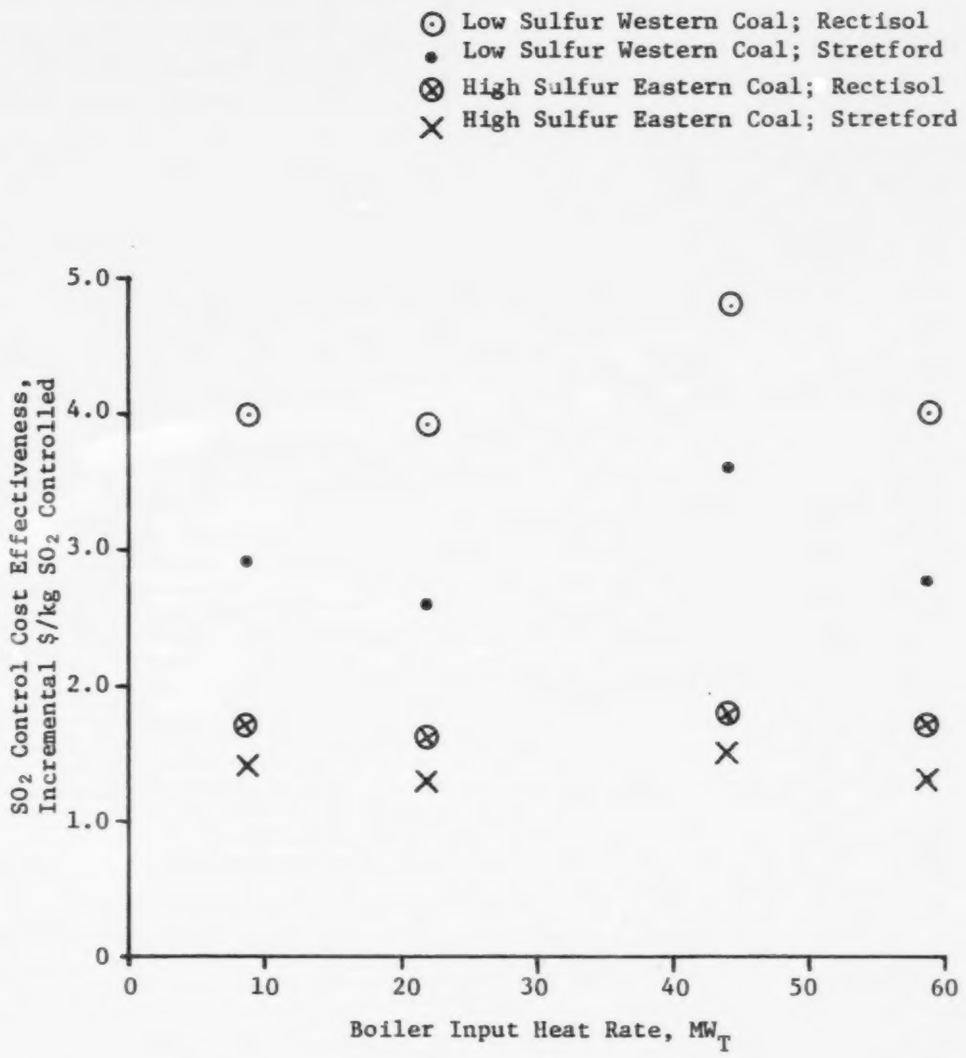


Figure 4.3-3. Cost effectiveness of medium-Btu gasification/steam generation system in controlling SO_2 emissions.

- Figure 4.3-1 presents the annualized medium-Btu gas-fired boiler costs as a function of boiler capacity.
- Figure 4.3-2 illustrates the incremental costs of emissions control (compared to uncontrolled coal-fired boilers) as a function of boiler size.
- Figure 4.3-3 graphically illustrates the cost effectiveness of medium-Btu gasification in controlling SO_2 emission assuming all costs of control are allocated to SO_2 emissions control.

Before discussing the bases used to calculate the cost data in Table 4.3-1 and Figures 4.3-1, 4.3-2, and 4.3-3, some comments are required concerning the use of the cost data and the need to consider both coal and medium-BTU gas transportation costs. First, the size and complexity of the medium-Btu gasification facilities prohibited detailed cost estimating in this study. The true costs for many of the pieces of equipment depend highly on factors which are specific, and sometimes unique, for each application. Many assumptions had to be made to develop the design and operating parameters which were required for the cost analyses. Thus, when using the cost estimates, one should be aware of the assumptions and bases that went into their development.

The second item is a consideration of the impact transportation costs have on the cost of the medium-Btu gas and hence on the annualized cost of the boiler. The coal costs used in this study are minemouth costs, as provided by the EPA and PEDCo (Ref. 4-5). They do not include transportation costs which can, in the case of western coals, increase delivered coal costs to as high as 2 or 3 times the minemouth costs. For the western coal cases, a \$1.00/GJ or 250 percent increase in coal costs would increase the medium-Btu gas costs by about \$1.70/GJ. That translates into a 40 percent increase in gas costs for the Stretford case and a 34 percent increase for the Rectisol case. For the eastern coal cases, a \$0.50/GJ or 74 percent increase in coal costs would cause the medium-Btu gas price to increase by \$0.98/GJ. That is a 16 percent increase in gas costs for the Stretford case and a 14 percent increase for the Rectisol case.

The medium-Btu gasification plants in this study are also assumed to be located away from, but within the general region of the industrial boiler. The cost of transporting the medium-Btu gas from the gasification facility to the boiler, however, has not been included in the cost analysis. This cost will obviously depend on the distance between the two locations, the volume of gas transmitted, and site specific factors such as the cost of procuring right-of-way for the transmission pipeline.

4.3.2 Cost Bases

The cost data presented in Table 4.3-1 were developed by first estimating the annualized costs of producing medium-Btu gas, and from that, the unit cost of the gas. The unit gas costs were used to calculate the annual fuel costs for the four sizes of industrial boilers examined. These were then incorporated into the annualized cost analysis of each boiler.

Estimates of the capital requirements and operating and maintenance costs for the four gasification systems were developed from data presented in a study by C. F. Braun for high-Btu gasification systems (Ref. 4-23). Adjustments were made to the C. F. Braun study data to 1) reflect the differences between high-Btu and medium-Btu gasification systems (deletion of the costs for the shift conversion, methanation and product gas compression units) and 2) incorporate the use of different operating conditions (coal, oxygen and steam flow rates) and acid gas/sulfur recovery processes. The size and cost of the on-site coal-fired utility boilers were also adjusted to reflect different steam and electric power requirements and to include the use of limestone flue gas desulfurization (FGD) units to control SO_2 emissions from the boiler.

The process areas for which capital requirements were estimated are shown in Table 4.3-2, along with the operating parameter used to adjust the C. F.

Braun data. For all adjustments except for the Rectisol costs for the high sulfur coal case, a scale factor of 0.6 was used to reflect the influence of economy of scale. As shown in Table 4.3-2, costs for several of the process areas were not adjusted. This was because, either 1) their costs were relatively minor and/or 2) sufficient detailed information was not available to permit making adjustments.

TABLE 4.3-2. CAPITAL COST ITEMS FOR MEDIUM-BTU GASIFICATION SYSTEMS

	Adjustment Parameter for C.F. Braun Costs
Coal Handling and Reclaiming	Coal Flow Rate
Coal Preparation	Coal Flow Rate
Coal Feed, Gasification and Raw Gas Quench	Not Adjusted
Acid Gas Removal	(1) (2)
Product Gas Drying	Not Adjusted
Liquid Effluent Treatment	Not Adjusted
Sulfur Recovery	(1)
Solids Disposal	Solids Flow Rate
Steam and Utilities Systems	Heat Duty and Electricity Required (3)
Plant Water Systems	Not Adjusted
Oxygen Plant	Oxygen Flow Rate
General Facilities	Not Adjusted

(1) Costs for the Stretford process used for acid gas removal/sulfur recovery and for just sulfur recovery were supplied by vendor quotes (Ref. 4-8).

(2) Costs could not be obtained for the Rectisol process. At the suggestion of Lurgi (Ref. 4-24), the licensor of the Rectisol process, costs presented by C.F. Braun for the Selexol process were used for the Rectisol costs in the low sulfur coal case. These costs were ratioed by the quantity of acid gas removed to estimate costs for the high sulfur eastern coal case (Ref. 4-25).

(3) Limestone FGD costs were estimated from Ref. 4-26 based on the quantities of flue gas to be treated and SO_2 removed.

The costs for the coal feed, gasification, raw gas quench and shift conversion units of the C. F. Braun study were provided by Lurgi and reported as a group. The cost of the shift conversion unit was estimated at $52 million (January 1976 dollars) based on cost data for other systems presented in the C. F. Braun study. Insufficient data were available to adjust the cost of the coal feed, gasification, and raw gas quench units.

Cost estimates for the Stretford process were obtained from process vendors (Ref. 4-8). Estimates for the Rectisol process could not be obtained. Instead, at the suggestion of Lurgi (the licensor of the Rectisol process), C. F. Braun's cost data for the Selexol process were used for the Rectisol costs in the low sulfur coal case (Ref. 4-24). For the high sulfur coal case, the Selexol costs were ratioed by the amount of acid gases removed. This is a reasonable approach for situations where most of the acid gas is CO_2 as is the case in this study (Ref. 4-25).

The costs of the limestone FGD unit for the utility boiler were estimated from data presented in Ref. 4-26. These costs were adjusted according to flue gas flow rate and quantity of SO_2 removed.

All estimates of capital requirements were indexed to mid-1978 dollars using the CE Plant Cost Index (mid-1978 = 217.7). Tables 4.3-3 and 4.3-4 summarize the adjusted capital investment requirements for the four medium-Btu gasification systems.

Operating costs for the gasification systems were also estimated based on data presented in C. F. Braun's study (Ref. 4-23). Maintenance costs and annualized operating costs for the four systems examined are summarized in Appendix B. Tables 4.3-5 and 4.3-6 for the low-sulfur western coal case using the Stretford process are shown here to indicate the cost approach taken and the bases used. Unit costs for the direct operating costs, e.g.,

TABLE 4.3-3. ESTIMATED INSTALLED EQUIPMENT COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS[a]

Process Area	Low Sulfur Western Coal		High Sulfur Eastern Coal	
	Stretford	Rectisol	Stretford	Rectisol
Coal Storage and Reclaiming	17	17	16	16
Coal Preparation	20	20	19	19
Coal Feed Gasification Raw Gas Quench	162	162	162	162
Acid Gas Removal	11	136	33	169
Product Gas Drying	0.9	0.9	0.9	0.9
Sour Water Stripping, Ammonia Recovery, and Bio-Oxidation	22	22	22	22
Sulfur Recovery	-	16	-	40
Solids Disposal	3.5	3.5	5.6	5.5
Steam and Utility Systems	228	233	324	316
Plant Water Systems	26	26	26	26
Oxygen Plant	77	77	95	95
General Facilities	96	96	96	96
TOTAL INSTALLED EQUIPMENT COSTS	663.4	809.4	799.5	967.4

[a] 3060 MW_T (250×10^9 Btu/day) capacity (gas output); mid-1978 dollars; 10^6 dollars.

TABLE 4.3-4. ESTIMATED TOTAL CAPITAL REQUIREMENTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS[a]

	Low Sulfur western coal		High sulfur eastern coal	
	Stretford	Rectisol	Stretford	Rectisol
Total Engineered Equipment Costs (EC) (including direct installation costs)	663.4	809.4	799.5	967.4
Installation Costs, indirect				
Engineering and Fee (10% of EC)	66.3	80.9	80.0	96.7
Construction and Field Expense (included in Equipment Costs)				
Contingencies (30% of EC)	218.9	267.1	263.9	319.2
Allowance for Funds Used During Construction (EC x Average Spending Period (1.75 yrs) x 10%)	116.1	141.6	139.9	169.3
Paid-up Royalties	1.1	1.1	1.1	1.1
Startup Costs (20% of total gross annual operating costs)	23.1	24.4	38.4	40.8
Land	0.9	0.9	0.9	0.9
Working Capital (25% total direct operating costs)	26.0	27.7	45.0	47.3
Total Capital Requirement	1115.8	1353.1	1368.7	1642.7

[a] 3060 MW_T (250×10^9 Btu/day) capacity (gas output); mid-1978 dollars; 10^6 dollars.

TABLE 4.3-5. ESTIMATED MAINTENANCE COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock	Low sulfur western
Acid Gas Removal Unit	Stretford

Unit	Unit Cost	Maintenance Factor, %	Maintenance Cost
Coal Handling and Reclaiming	24	6	1.4
Coal Preparation	29	6	1.7
Coal Feed Gasification Raw Gas Quench	232	5	11.6
Acid Gas Removal	16	3	0.5
Product Gas Drying	1	3	0.03
Liquid Effluent Treatment	31	3	0.9
Sulfur Recovery	-	3	-
Solids Disposal	5	3	0.2
Steam and Utilities Systems	326	1	3.3
Plant Water Systems	37	3	1.1
Oxygen Plant	110	3	3.3
General Facilities	137	1	1.4
TOTAL	948		25.4
Maintenance Labor (@ 60% of Total Maintenance)			15.2
Maintenance Supplies (@ 40% of Total Maintenance)			10.8

[a] 3060 MW_T (250×10^9 Btu/day) gas output capacity; 10^6 mid-1978 dollars.

TABLE 4.3-6. ANNUALIZED COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
Operating Factor - 90%

Direct Costs			
Operating Labor (@ \$12.02/hr)	6.3		
Maintenance Labor (60% of Total Maintenance)	15.2		
Supervision (20% of Operating and Maintenance Labor)	4.3		
Maintenance Materials (40% of Total Maintenance	10.8		
Replacement Parts (Included in Maintenance Materials)	-		
Operating Supplies (30% of Operating Labor)	1.9		
Purchased Water (@ \$0.032/m^3)	0.1		
Coal (@ \$8.88/ton)	58.3		
Chemicals and Catalysts	2.3		
Solids Disposal (@ \$11/ton)	4.7		
Total Direct Costs		103.9	
Overhead Costs			
Payroll (30% of Operating Labor)	1.9		
Plant (26% of Labor, Materials, and Maintenance)	9.5		
Total Overhead Costs		11.4	
By-Product Credits			
Sulfur (@ \$29/ton)	0.8		
Ammonia (@ \$110/ton)	5.2		
Naphtha (@ \$93/m^3)	-		
Phenols (@ \$46/m^3)	1.1		
Tars (@ \$46/m^3)	8.7		
Oils (@ \$79/m^3)	8.7		
Total By-Product Credits		(24.5)	
Capital Related Charges			
G&A, Local Taxes, and Insurance (4% of Total Capital Requirement excluding Start-up Costs, Land, and Working Capital)	42.6		
Capital Recovery (11.75% of Total Capital Requirement excluding Land and Working Capital)	127.9		
Interest on Working Capital (@ 10%)	2.6		
Total Capital Related Charges		173.1	
TOTAL ANNUALIZED COSTS			263.9
Average Gas Costs, \$/GJ (\$/10^6 Btu)			3.05 (3.20)
Average Gas Costs[b], \$/GJ (\$/10^6 Btu)			4.30 (4.55)

[a] 3060 MW_T (250 x 10^9 Btu/day) output gas capacity; 10^5 mid-1978 dollars.

[b] At 60% operating factor.

coal costs, were those indicated in Table 4.1-1 (Ref. 4-5). The average gas costs shown in Table 4.3-6 were calculated in a manner analogous to that described previously for low-Btu gasification systems. The following assumptions were used:

- 20-year project life,
- 10 percent interest rate, and
- 60 and 90 percent operating factors.

The medium-Btu gas plant is designed to operate at a 90% operating factor. Average gas costs were calculated based on this operating factor. However, an industrial boiler with a 60% operating factor will have to pay more for gas from this facility. There are various means available to permit the centralized gas plant to operate at full capacity 90% of the time. They include:

- Gas storage capability could be made available
- Backup fuels could be utilized by industrial boiler operators during peak requirements.

These techniques were not evaluated because of the additional complexity which would be added to the study. Therefore medium-Btu gas costs were also evaluated using a 60% operating factor. The costs (at 60% operating factor) were used in Table 4.3-1 in developing annualized medium-Btu gas-fired boiler costs.

Annualized operating costs for the medium-Btu gas-fired boilers were also estimated in a manner identical to that for the low-Btu gas-fired boilers. Capital requirements and all direct operating costs except fuel costs were assumed to be the same as for the same capacity low-Btu gas-fired boiler. Annualized capital charges were calculated using a capital recovery factor based on a 20-year boiler lifetime and interest at 10 percent. Detailed summaries of the annualized costs of the 16 medium-Btu gas-fired boilers examined are contained in Appendix B.

4.3.3 Cost Sensitivity Analysis

In this section, the major cost components of the medium-Btu gas-fired boiler annualized costs are identified. Specifically the effect of coal costs on annualized boiler costs is evaluated since the base case coal costs are typical of minemouth costs and some degree of coal handling and transportation is likely for most industrial application.

A complete breakdown of the annualized costs for medium-Btu gas-fired boilers is presented in Table 4.3-7.

Since the coal-to-gas conversion efficiency is less than unity, an increase in coal cost will increase the annualized costs of a medium-Btu gas-fired boiler relative to those of an uncontrolled coal-fired boiler. Examples are presented in Table 4.3-8. For illustrative purposes, coal cost increases of $1.00/GJ and $0.50/GJ are examined for western and eastern coal, respectively. These are representative of possible coal transportation costs expected for coals in these regions.

4.4 COAL LIQUEFACTION

The SRC-I and Exxon Donor Solvent (EDS) processes were selected as the "best candidate" coal liquefaction systems for controlling emissions from industrial boilers. Capital requirements for constructing facilities using these processes and annualized operating costs were estimated from available literature data (Ref. 4-27, 4-28). The coal feedstocks considered in the cost estimates were high sulfur bituminous coals. Insufficient information was available to develop cost estimates for a low sulfur coal feedstock.

From the annualized cost estimates for the coal liquefaction facilities, unit fuel prices were calculated. These were incorporated into cost analyses

TABLE 4.3-7. SENSITIVITY ANALYSIS - MEDIUM-BTU GAS-FIRED BOILER ANNUALIZED COSTS

Coal Boiler Input Heat Rate, MW_T	Low Sulfur Western		High Sulfur Eastern	
	8.8	58.6	8.8	58.6
	Percent of Annualized Costs			
Labor and Maintenance	19.8	7.0	16.1	5.3
Utilities	2.5	1.0	2.0	0.6
Overhead	7.4	2.9	6.0	2.2
Capital Charges	10.7	7.4	8.7	6.2
Fuel (Medium-Btu Gas)				
Labor & Maintenance	8.7	12.0	7.6	9.7
Coal	9.2	12.7	15.0	19.2
Other Operating Costs	1.5	2.1	3.2	4.1
By Product Credits	(3.9)	(5.3)	(1.9)	(2.5)
Overhead	2.7	3.8	2.3	3.0
Capital Charges	41.3	56.5	40.9	52.1
Total Fuel	59.5	81.7	67.1	85.6
Annualized Costs, 10^3\$/yr	1210	5840	1490	7770

Basis: 1) Stretford is the acid gas removal unit for all cases shown

2) Operating factors - 60% for gasification plant
60% for boiler

3) The first four items in the table above - labor and maintenance, utilities, overhead, capital charges - refer to the gas-fired boiler system. The fifth item - fuel - is medium-Btu gas from the gasification plant. Percentages under fuel are the various components of the gas costs.

TABLE 4.3-8. EFFECT OF COAL TRANSPORTATION COSTS ON ANNUALIZED MEDIUM-BTU GAS-FIRED BOILER COSTS

Input Heat Rate to Boiler, MW_T (10^6 Btu/hr)	Coal Cost = Base Case Values[a]			Coal Transportation Costs Added[b]		
	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Medium-Btu gas-fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr	Coal-fired Boiler Annualized Costs, $\$10^3$/yr	Medium-Btu gas-fired Boiler Annualized Costs, $\$10^3$/yr	Incremental Costs, $\$10^3$/yr
Low Sulfur Western Coal						
8.8(30)[c]	977	1210	230	1144	1490	350
8.8(30)[d]	977	1320	340	1144	1600	460
58.6(200)[c]	4369	5840	1470	5478	7720	2240
58.6(200)[d]	4369	6630	2260	5478	8520	3040
High Sulfur Eastern Coal						
8.8(30)[c]	952	1490	540	1035	1640	610
8.8(30)[d]	952	1650	700	1035	1800	770
58.6(200)[c]	4248	7770	3520	4802	8800	4000
58.6(200)[d]	4248	8790	4540	4802	9820	5020

[a] Base Case Values: Low Sulfur Western (LSW) - $0.40/GJ; High Sulfur Eastern (HSE) - $0.68/GJ

[b] Transportation Costs: LSW - $1.00/GJ; HSE - $0.50/GJ

[c] Gasification System Uses Stretford Process

[d] Gasification System Uses Rectisol Process

developed for several sizes of industrial boilers. For the EDS liquids, four package, watertube boilers sizes were examined--8.8, 22, 44, and 58.6 MW_T (30, 75, 150, and 200 x 10^6 Btu/hr) heat input. For SRC-1, the only boiler considered was a 58.6 MW_T (200 x 10^6 Btu/hr), field-erected pulverized coal type boiler. Smaller boilers were not examined because they generally would be stoker types, and it is impractical to use SRC-I in a stoker type boiler due to its low melting point.

The estimated annualized costs for industrial boilers firing EDS liquids and SRC-I are summarized in Table 4.4-1. Also included in this table are the percentage increases that these annualized costs represent over the costs of direct coal-fired boilers (Ref. 4-3).

Emissions from the combustion of SRC-I and EDS liquids are summarized in Table 4.4-2. Also shown in this table are the target emission control levels for combustion of coal-derived liquids. As indicated in this table, SRC-I can only meet the moderate SO_2 emission control level, while raw EDS liquids can meet the intermediate level, and if hydroprocessed, can meet the stringent level. If a 80^+ percent efficient particulate control device is used on the boiler flue gases, SRC-I combustion can meet the particulate emission control target of 13 ng/J (0.03 lb/10^6 Btu). The EDS liquids can meet the target control level without a post-combustion particulate control device. NO_x emissions indicated in Table 4.4-2 are based on EPA emission factors for coal and residual fuel (Ref. 4-29). Whether these factors are appropriate for coal-derived fuels is not known. Combustion modification potentially could reduce the levels of NO_x emissions. However, the extent to which combustion modifications would be effective is not known.

4.4.1 Cost Bases

Annualized costs for the SRC-I and EDS facilities were developed from available literature data (Ref. 4-27, 4-28). Capital requirements (indexed to mid-1978 dollars using the M+S cost index) for the facilities are

TABLE 4.4-1. COST ANALYSIS OF "BEST CANDIDATE" COAL LIQUEFACTION SYSTEMS FOR INDUSTRIAL BOILERS

Boiler Heat Input, MW_T (10^6 Btu/hr)	Coal Feed to Liquefaction Process	Coal Liquefaction Process	Emission Control Efficiency, %		Annualized Coal Liquids-Fired Boiler Costs		% Increase in Costs Over Uncontrolled Coal-Fired Boiler Costs
			SO_2	Particulates	10^3 \$/yr	\$/kW_T (\$/10^6 Btu/hr)	
8.8 (30)	High Sulfur Bituminous	EDS	95*	99.4	1,300	148 (43,000)	37
22 (75)	High Sulfur Bituminous	EDS	95*	99.4	2,670	121 (36,000)	44
44 (150)	High Sulfur Bituminous	EDS	95*	99.8	5,110	116 (34,000)	66
58.6 (200)	High Sulfur Bituminous	EDS	95*	99.8	6,580	112 (33,000)	55
58.6 (200)	High Sulfur Bituminous	SRC-I	83	99.6**	6,800	116 (34,000)	60

* Additional hydroprocessing can increase SO_2 control efficiency to 96.5% or greater.

** Assumes 80% efficiency particulate control device on boiler flue gases.

All costs are mid-1978 dollars.

TABLE 4.4-2. EMISSIONS FROM COAL LIQUIDS COMBUSTION

	SO_2, ng/J	NO_x, ng (as NO_2)/J	Particulates, ng/J
SRC-I*	440	(248)	13**
EDS Liquids*			
Raw	125	(115)	6
Hydroprocessed	83	(115)	6
Target Control Levels			
Moderate Control	520	300	13
Intermediate Control	260	200	13
Stringent Control	86	86	13

* Produced from high sulfur bituminous coal.

** Assumes 80% efficient particulate control device on flue gas.

Numbers in parentheses are merely estimates based on EPA emissions factors (Ref. 4-29). The validity of applying those emission factors to coal liquids combustion is not known.

presented in Tables 4.4-3 and 4.4-4. Annualized costs and average product costs are presented in Tables 4.4-5 and 4.4-6. The following assumptions were used to develop the cost estimates shown in these tables:

- Labor, process water, catalyst, and chemical costs for the SRC-I facility were assumed to be equal to those for the EDS facility.
- Minemouth coal costs were used, i.e., costs for transporting the coal feedstock from the mine to the liquefaction facility were not included.
- A 20-year equipment life and 10 percent annual interest rate were used to calculate the annualized capital charges.
- An 80 percent plant operating factor was used to calculate the average product costs.

Annualized costs for the coal liquids-fired boilers were estimated from data developed by PEDCo (Ref. 4-3) for a 58.6 MW_T, pulverized low sulfur eastern field-erected coal-fired boiler (for the SRC-I boiler) and for a 44 MW_T residual fuel oil-fired package boiler (for the EDS liquids-fired boilers). Capital requirements for the 8.8, 22, and 58.6 MW_T EDS liquids-fired boilers were estimated using the 44 MW_T boiler capital requirement. The assumptions used in developing the annualized boiler costs included:

- The cost of transporting the coal liquids from the coal liquefaction plant to the industrial boiler were not included.
- A 45-year equipment life and 10 percent annual interest rate were used to calculate the annualized capital charges for the SRC-I-fired boiler.
- A 20-year equipment life and 10 percent annual interest rate were used to calculate the annualized capital charges for the EDS liquids-fired boilers.
- A 60 percent annual operating factor was assumed for all of the boilers.

Detailed summaries of the capital requirements and annualized costs of the SRC-I and EDS liquids-fired boilers are contained in Appendix C.

TABLE 4.4-3. TOTAL CAPITAL REQUIREMENT FOR SRC-I PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	7050 MW_T (577 x 10^9 Btu/day)
SO_2 Control Level	Moderate
Installed Equipment Costs	
Coal preparation	90
Hydrogenation and hydrogen recycle	221
Fractionation	27
Hydrogen plant	342
Filtration	192
Product solidification	28
Gas and secondary recovery	90
Offsites and wastewater treatment	110
General Facilities	86
Total Installed Equipment (EC)	1,186
Engineering and fee (included in direct costs)	-
Construction and field expense (included in direct costs)	-
Contingencies (30% of EC)	356
Start-up Costs (20% of total gross annual operating costs)	59
Allowance for funds used during construction (EC x 1.75 x 0.10)	208
Land	2
Working capital (25% of total direct annual operating costs)	73
Total Indirect Costs	698
TOTAL CAPITAL REQUIREMENT	1,884

10^6 mid-1978 dollars

TABLE 4.4-4. TOTAL CAPITAL REQUIREMENTS FOR EXXON DONOR SOLVENT PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	4000 MW_T (328 x 10^9 Btu/day)
SO_2 Control Level	Intermediate*
Installed Equipment Costs	
Liquefaction	218
Solvent hydrogenation	74
Flexicoker	145
Hydrogen recovery and generation	218
Gas and wastewater treatment	44
Product recovery	7
Offsites	224
Total Equipment Costs (EC)	930
Engineering and fee (10% of EC)	93
Construction and field expense (included in direct costs)	
Contingencies (30% of EC + Engineering + Fee)	307
Start-up costs (20% of total gross annual operating costs	53
Allowance for funds used during construction (EC x 1.75 x 0.10)	163
Land	2
Working capital (25% of total direct annual operating costs)	66
Total Indirect Costs	684
TOTAL CAPITAL REQUIREMENT	1,614

*Additional hydroprocessing will permit attainment of stringent control levels. Additional costs incurred for hydroprocessing cannot be determined and are assumed to be negligible.

10^6 mid-1978 dollars

TABLE 4.4-5. ANNUALIZED COSTS FOR SRC-I PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	7050 MW_T (577 x 10^9 Btu/day)
SO_2 Control Level	Moderate
Operating Factor	80%
Direct Costs	
Manpower	31
Repair materials and other	48
Electricity (@ 25.8 mills/kW-h)	16
Purchased water (@ $0.032/$m^3$)	<1
Catalysts and Chemicals	6
Coal (@ $18.72/ton)	163
Total Direct Costs	264
Overhead	
Payroll (30% of labor)	9
Plant (26% of labor, parts and maintenance)	20
Total Overhead Costs	29
General and Administrative Costs, Local Taxes and Insurance (4% of total capital requirement excluding land and working capital)	72
By-Product Credits	(9)
Capital Recovery Factor (11.75% of total capital requirement excluding land and working capital)	213
Interest on working capital	7
TOTAL ANNUALIZED COSTS	576
Average Product Cost, $/GJ	3.25

10^6 mid-1978 dollars

TABLE 4.4-6. ANNUALIZED COSTS FOR EXXON DONOR SOLVENT PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	4000 MW (328 x 10^9 Btu/day)
SO_2 Control Level	Intermediate*
Operating Factor	80%

Direct Costs	
Manpower	31
Repair materials and other	48
Electricity (@ 25.8 mills/kW-h)	32
Purchased water (@ $0.032/$m^3$)	<1
Catalysts and Chemicals	6
Coal (@ $18.72/ton)	119
Total Direct Costs	236
Overhead	
Payroll (30% of labor)	9
Plant (26% of labor, parts and maintenance)	20
Total Overhead Costs	29
General and Administrative Costs, Local Taxes and Insurance (4% of total capital requirements excluding land and working capital)	62
By-Product Credits	(16)
Capital Recovery Factor (11.75% of total capital requirement excluding land and working capital)	213
Interest on working capital	7
TOTAL ANNUALIZED COSTS	576
Average Product Cost, $/GJ	4.95

*Additional hydroprocessing will permit attainment of stringent control levels. Additional costs incurred for hydroprocessing cannot be determined and are assumed to be negligible.

10^6 mid-1978 dollars

4.4.2 Costs Summary and Analysis

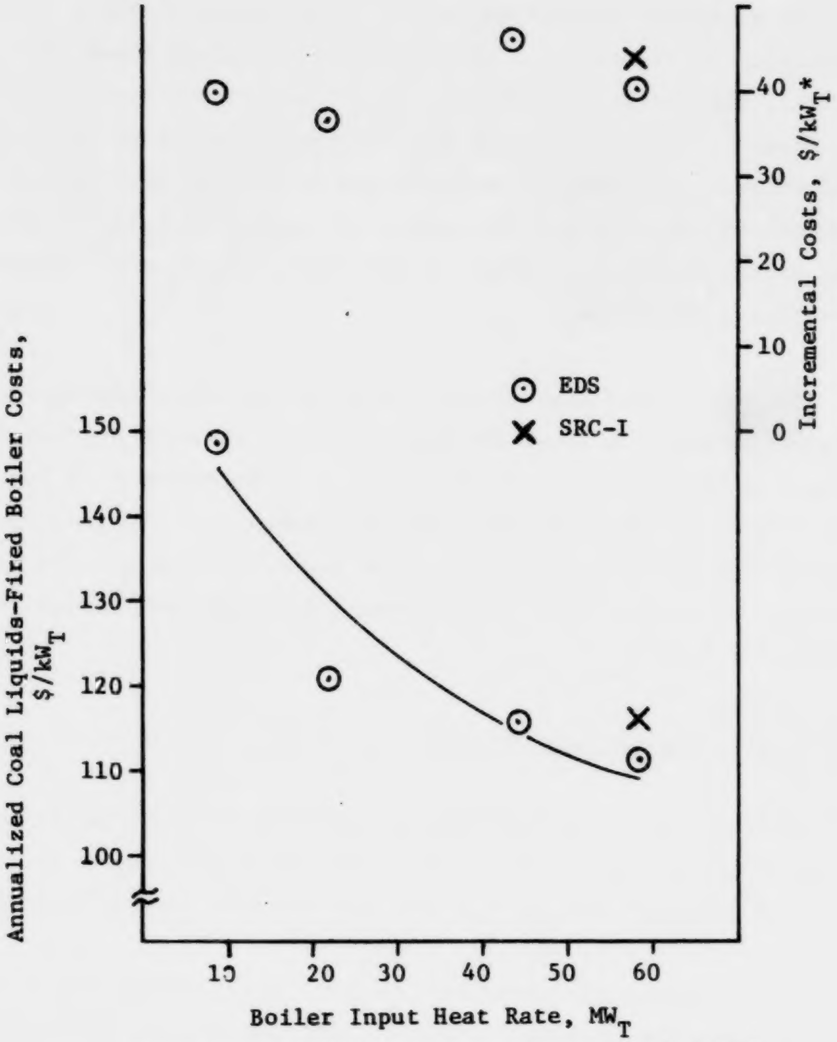
The estimated annualized boiler costs listed in Table 4.4-1 are shown graphically in Figure 4.4-1 as a function of boiler input heat rate. As expected, the larger boilers have lower annualized costs per unit of capacity. Also shown in Figure 4.4-1 are the incremental costs of the coal liquids-fired boilers (in $\$/kW_T$ of design input heat rate) over direct coal-fired boilers. As was true for the medium-Btu gas-fired boilers, the coal liquids-fired boiler incremental costs do not have a significant dependence on boiler input heat rate.

The SO_2 control cost effectiveness of the coal liquefaction systems as emission controls is shown in Figure 4.4-2. The data in this figure are expressed as incremental costs per kg of SO_2 controlled as a function of boiler size. This assumes that the total incremental cost of control for coal liquefaction systems are allocated to SO_2 control only. The figure indicates that the SO_2 control cost effectiveness does not have a significant dependence on boiler capacity.

4.4.3 Cost Sensitivity Analysis

In Table 4.4-7 the individual components of the coal liquids-fired boiler annualized costs are identified. This analysis is for the 58.6 MW_T SRC-I-fired boiler and the 8.8 and 58.6 MW_T EDS liquids-fired boilers.

The cost data presented for the coal liquefaction facilities and the coal liquids-fired boilers do not include the cost of transporting coal from the mine to the coal liquefaction facility or transporting the coal liquids to the industrial boiler. In order to provide insight into the impact that transportation costs could have, a rough cost sensitivity analysis was performed. Two cases were examined for the 58.6 MW_T SRC-I-fired boiler and the 8.8 and 58.6 MW_T EDS liquids-fired boilers:



*Incremental dollars above direct coal-fired boiler costs.

Figure 4.4-1. Cost analysis of coal liquids-fired industrial boilers.

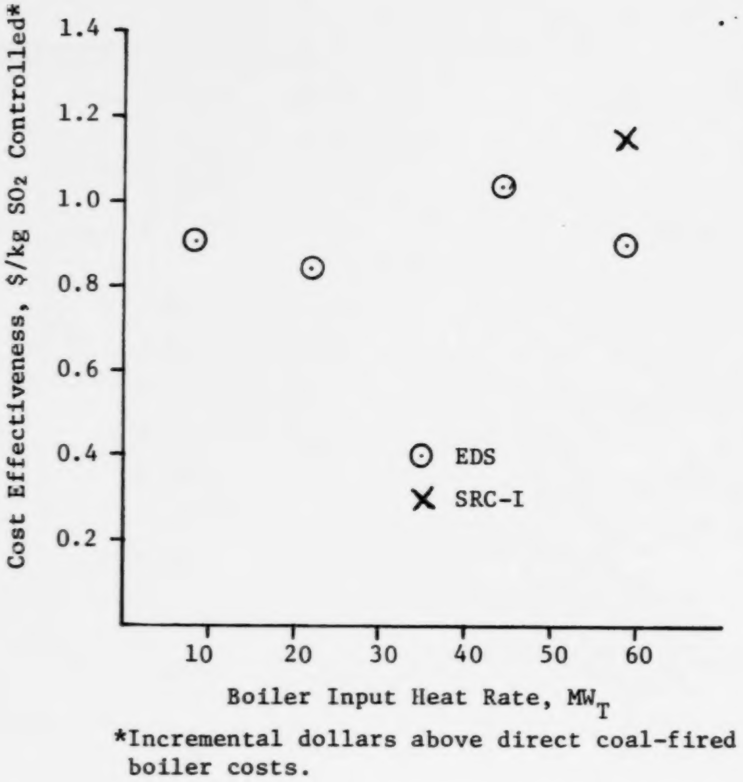


Figure 4.4-2. Cost effectiveness of coal liquefaction systems as SO_2 emission controls for industrial boilers.

TABLE 4.4-7 COST SENSITIVITY - COAL LIQUIDS-FIRED BOILERS

Coal Liquefaction Process	EDS		SRC-1
Boiler Input Heat Rate, MW_T	8.8	58.6	58.6
	Percent of Annualized Costs		
Labor and Maintenance	18.0	6.3	14.7
Utilities	2.4	0.9	3.7
Overhead	7.1	2.6	5.7
Capital Charges	9.1	6.8	23.0
Fuel			
Labor & Maintenance	10.1	13.2	7.2
Coal	15.1	19.8	15.0
Other Operating Costs	2.8	3.7	1.2
Overhead	3.7	4.8	2.6
Capital Charges	31.7	41.9	26.9
Total Fuel	63.4	83.4	52.9
Total	100.0	100.0	100.0
Total Annualized Boiler Costs, 10^3\$/yr	1300	6580	6800

1) Base case coal costs (\$18.72/ton or \$0.68/GJ) with \$0.50/GJ coal transportation costs added.

2) Minemouth coal costing \$0.68/GJ with \$0.50/GJ coal transportation and liquefaction plant product transportation of \$0.50/GJ.

In order to make valid comparisons, annualized costs were also calculated for direct coal-fired boilers using \$1.18/GJ coal.

The results of the sensitivity analysis are presented in Table 4.4-8, and illustrate the significant impact that transportation costs can have. For example, the base case incremental costs for the 8.8 MW EDS liquids-fired boiler (\$18.72/ton coal, no transportation costs) are \$40/kW_T. However, for the worst case examined, i.e., \$1.18/GJ coal and \$0.50/GJ product fuel transportation costs, the incremental costs increase about 40 percent to \$56/kW_T.

4.5 SUMMARY

Table 4.5-1 presents summary information concerning the three synfuel technologies evaluated. All of the synfuel plant configurations are not described. For example, the SRC-I-fired boiler data are not presented. Also, only data for two boiler heat input rates (8.8 and 58.6 MW_T) are presented. The configurations presented are not intended to be "best cases" but rather were selected for illustrative purposes.

In all of the systems studied, the capital-related charges and O&M costs are the largest contributors to annualized costs. The coal cost is the next largest contributor. Since a minemouth coal cost was used in the study, the coal component could be much larger when a realistic delivered cost (including coal transportation) is included in the analysis.

A brief characterization of each synfuel technology follows:

TABLE 4.4-8. EFFECT OF COAL AND PRODUCT TRANSPORTATION COST ON ANNUALIZED COAL LIQUIDS-FIRED BOILER COSTS

Boiler Heat Input, MW_T	Process	Coal Cost $/GJ	Coal Liquid Cost, $/GJ	Product Transportation Cost, $/GJ	Liquids Delivered Cost, $/GJ	Annualized Synfuel-Fired Boiler Costs, 10^3\$/year	Annualized Coal-fired Boiler Costs, 10^3\$/yr	Incremental Cost $/year	$/KW_T$
8.8	EDS	0.68	4.95	--	4.95	1300	950	350	40
8.8	EDS	1.18	5.82	--	5.82	1450	1040	410	46
8.8	EDS	1.18	5.82	0.50	6.32	1530	1040	490	56
58.6	EDS	0.68	4.95	--	4.95	6580	4250	2330	40
58.6	EDS	1.18	5.82	--	5.82	7550	4820	2730	47
58.6	EDS	1.18	5.82	0.50	6.32	8100	4820	3280	56
58.6	SRC-1	0.68	3.25	--	3.25	6800	4250	2550	44
58.6	SRC-1	1.18	3.93	--	3.93	7560	4820	2740	47
58.6	SRC-1	1.18	3.93	0.50	4.43	8110	4820	3290	56

TABLE 4.5-1. SYNFUEL ECONOMICS SUMMARY

	Percent of Total Annualized Boiler Costs[a]							Total Annualized Boiler Costs, 10^3 $/year	Incremental Annualized Boiler Costs (over uncontrolled coal-fired boiler costs)	
				Fuel						
	Labor and Maintenance[b]	Utilities[c]	Capital-Related Charges	Labor and Maintenance[b]	Coal	Other	Capital-Related Charges		10^3 $/year	$/kW$_T$
Low Sulfur Western Coal[d]										
Uncontrolled Coal-Fired Boiler										
8.8 MW$_T$	53.5	6.6	33.1	-	6.8	-	-	980	-	-
58.6 MW$_T$	34.9	8.6	46.4	-	10.1	-	-	4370	-	-
Low-Btu Gas-Fired Boiler[e]										
8.8 MW$_T$	15.7	1.4	5.2	32.8	4.3	4.3	36.3	2100	1120	127
58.6 MW$_T$	10.9	1.1	5.8	25.0	11.3	12.2	33.6	5320	950	16
Medium-Btu Gas-Fired Boiler[f]										
8.8 MW$_T$	27.2	2.5	10.7	11.4	9.2	(2.4)	41.3	1210	230	26
58.6 MW$_T$	9.9	1.0	7.4	15.8	12.7	(3.2)	56.5	5840	1470	25
High Sulfur Eastern Coal[d]										
Uncontrolled Coal-Fired Boiler										
8.8 MW$_T$	52.7	6.7	28.3	-	12.3	-	-	950	-	-
58.6 MW$_T$	33.9	7.7	40.1	-	18.3	-	-	4250	-	-
Low-Btu Gas-Fired Boiler[e]										
8.8 MW$_T$	14.2	1.3	4.7	31.0	6.9	6.9	34.9	2320	1370	156
58.6 MW$_T$	8.9	0.9	4.7	21.4	16.5	16.8	30.6	6530	2280	39
Medium-Btu Gas-Fired Boiler[f]										
8.8 MW$_T$	22.1	2.0	8.7	9.9	15.0	1.3	40.9	1490	540	61
58.6 MW$_T$	7.5	0.6	6.2	12.7	19.2	1.6	52.1	7770	3520	60
EDS Liquids-Fired Boiler										
8.8 MW$_T$	25.1	2.4	9.1	13.8	15.1	2.8	31.7	1300	350	40
58.6 MW$_T$	8.9	0.9	6.8	18.0	19.8	3.7	41.9	6580	2330	40

[a] The first three items refer to annualized cost associated with the operation of the boiler only. The fuel item can be either coal or synfuel and the various components are broken down accordingly.

[b] Including overhead.

[c] Includes utilities, solid waste disposal (where applicable), water, chemicals.

[d] Coal costs are $0.68/GJ and $0.40/GJ for high sulfur Eastern and low sulfur Western, respectively. These are representative of minemouth costs.

[e] Stretford acid gas removal system.

[f] Rectisol acid gas removal system.

Low-Btu Gasification - This control technology is more cost effective at the higher capacities evaluated. Higher coal costs (resulting from including coal transportation costs) will tend to increase the incremental annualized boiler costs (over uncontrolled coal-fired boiler costs).

Medium-Btu Gasification - This control technology exhibits fairly constant incremental costs ($\$/kW_T$) as a function of boiler size. It should be pointed out that coal and medium-Btu gas product transportation costs have not been included in the evaluation and may significantly increase the incremental costs.

Coal Liquefaction - This technology also exhibits fairly constant incremental costs ($\$/kW_T$) as a function of boiler size. Again, coal and liquids transportation costs have been excluded and could increase the incremental costs substantially.

References

4-1. Guthrie, Kenneth M. Process Plant Estimating Evaluation and Control. Solana Beach, CA. Craftsman Book Co. 1974.

4-2. Peters, Max S., and Klause D. Timmerhaus. Plant Design and Economics for Chemical Engineers. Second ed. New York, NY. McGraw-Hill. 1968.

4-3. Devitt, T., P. Spaite, and L. Gibbs. The Population and Characteristics of Industrial/Commercial Boilers. EPA Contract No. 68-02-2603, Task 19. Cincinnati, Ohio. PEDCo Environmental, Inc. May 1979.

4-4. Telephone communication between Radian Engineer W. C. Thomas and David Noe of PEDCo Environmental, Inc. concerning various cost items for gas- and coal-fired boilers. 31 May and 1 June 1979.

4-5. PEDCo Environmental Specialists, Inc. "Emission Control System Economics." Section 3.0. Cincinnati, OH. October 1978.

4-6. Cavanaugh, E. C., W. E. Corbett and G. C. Page. Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology. Volumes I & II. Final Report. EPA 600/7-77-125a and b, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. September 1977.

4-7 Telephone communication between Radian Engineer W. C. Thomas and Dale Williams of J. F. Pritchard and Co. concerning costs for Stretford Process. 30 May 1979.

4-8. Telephone communication between Radian Engineer P. J. Murin and Dale Williams and Buzz Zey of J. F. Pritchard & Company concerning costs of Stretford Systems. 30 October 1978.

4-9. Letter from Roger S. Leland of J. F. Pritchard & Company to Radian Engineer T. G. Sipes concerning Stretford process. 25 July 1978.

4-10. Telephone communications between Radian Engineer P. J. Murin and Tom Jones and "Hub" Wills of Perry Gas Company concerning costs of MEA acid gas removal processes. 20 October 1978 through 2 November 1978.

4-11. Perry, Charles R. "Basic Design and Cost Data on MEA Treating Units." In Proceedings of the 1967 Gas Conditioning Conference. University of Oklahoma. Norman, OK.

4-12. Telephone communication between Radian Engineer P. J. Murin and Ken Braden of Ralph M. Parsons Company concerning costs for Claus units and tail gas treatment. 31 October 1978.

4-13. Telephone communication between Radian Engineer P. J. Murin and Jack McNamera of Shell Development Company concerning costs for SCOT process. 26 October 1978.

4-14. Dravo Corporation. Handbook of Gasifiers and Gas Treatment Systems. Final Report. Task Assignment No. 4. Report No. FE-1772-11. ERDA Contract No. E(49-18)-1772. Pittsburg, PA. Chemical Plants Division. February 1976.

4-15. Goar, Gene. "Impure Feeds Cause Claus Plant Problems." In Hydrocarbon Process. 53(7), 129-32. 1974.

4-16. Beavon, David K., and Raoul P. Vaell. "The Beavon Sulfur Removal Process for Purifying Claus Plant Tail Gas." In American Petroleum Institute Proceedings, Division of Refining, 1972. New York, NY. p. 267. 1972.

4-17. Telephone communication between Radian Engineers W. C. Thomas and P. J. Murin and Wallace Hamilton of McDowell Wellman concerning cost of the Wellman-Galusha gasifier. 25 May 1979.

4-18. Telephone communication between Radian Engineer W. C. Thomas and Hugh Campbell of EBECO Associates, Inc. concerning costs of coal handling equipment for Wellman-Galusha gasifier. 29 May 1979

4-19. Telephone communication between Radian Engineer W. C. Thomas and Mike Zolanzes of Research Cottrell concerning costs of detarrers. 30 May 1979.

4-20. Environmental Protection Agency. Process Design Manual for Suspended Solids Removal. EPA 625/1-75-003a. January 1975.

4-21. Fair, James R. "Designing Direct-Contact Coolers/Condensers." In Chem. Eng. 79(13), 91. 1972.

4-22. Fair, James R. "Process Heat Transfer by Direct Fluid-Phase Contact." AIChE Symp. Ser. 68(118), 1-11. 1972.

4-23. Detman, Roger F. Factored Estimates for Western Coal Commercial Concepts. Interim Report. Report No. FE-2240-5. Series No. IV: B-1. ERDA Contract No. E(49-18)-2240. Alhambra, CA. C. F. Braun & Company. October 1976.

4-24. Telephone communication between Radian Engineer P. J. Murin and Ted Pollaert of American Lurgi Corporation concerning costs for Rectisol process. 11 November 1978.

4-25. Telephone communication between Radian Engineer W. C. Thomas and Kim Crawford of TRW concerning costs for Rectisol and Selexol processes. 13 November 1978.

4-26. Torstrick, R. L., L. J. Henson and S. V. Tomlinson. "Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization." Presented at the Flue Gas Desulfurization Symposium. Hollywood, FL. November 1977.

4-27. Schmid, B. K., and D. M. Jackson, Gulf Mineral Resources Co. Recycle SRC Processing for Liquid and Solid Fuels. Presented at the Fourth Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity, University of Pittsburgh, Pittsburgh, PA. August 2-4, 1977.

4-28. Exxon Research and Engineering Company. EDS Coal Liquefaction Process Development, Phase III A. Interim Report. January 1978.

4-29. Environmental Protection Agency. Compilation of Air Pollutant Emission Factors, 2nd ed., with supplements. AP-42. Research Triangle Park, NC. February 1972, April 1973, July 1974, January 1975, December 1975, February 1976, April 1977.

SECTION V

ENERGY IMPACTS OF SYNTHETIC FUELS FROM COAL SYSTEMS

"Best candidate" synthetic fuels from coal systems were selected in Section 3 for detailed analysis of costs, energy, and environmental impacts. The two low-Btu gasification systems selected were the Wellman-Galusha gasifier with the Stretford and MEA acid gas removal (AGR) processes, while the two medium-Btu systems were the Lurgi gasifier with the Stretford and Rectisol AGR processes. The SRC-I and EDS systems were selected as the "best candidate" coal liquefaction systems. The results of the cost impact analyses were presented in Section 4.

Presented in this section are the energy impacts associated with using the "best candidate" systems as precombustion emission controls for new industrial boilers. Low-Btu coal gasification is addressed in Section 5.1 while Section 5.2 deals with medium-Btu gasification. Coal liquefaction is addressed in Section 5.3.

The largest consumption of energy in the synthetic fuels from coal processes is the inefficiency in the conversion process. Only 60-75 percent of the energy content of the coal feed to the gasifier or liquefaction reactor is converted into energy in the plant's major products.

In addition to the conversion inefficiency, the synthetic fuels processes consume energy in the form of steam to drive pumps, compressors, fans, etc. and to provide hydrogen for the liquefaction and gasification reactions. Electricity is also used to drive pumps, fans, agitators, etc., as well as to satisfy general lighting, cooling and heating needs. The plant's steam and electricity requirements can be supplied by on-site boilers or by importing these auxiliaries from an off-site source.

As mentioned previously, Standard International units are used in this report. Therefore, megawatts and kilowatts are used for both thermal power (denoted MW_T or kW_T) and electrical power (denoted MW_e or kW_e).

5.1 LOW-BTU COAL GASIFICATION

Energy consumption data for the low-Btu Wellman-Galusha gasifier and its associated gas clean-up systems (see Figure 5.1-1) are given in Table 5.1-1. The energy consumed by a low-Btu gasification system is linearly proportional to the plant's gas production rate. Therefore, in Table 5.1-1, only a boiler heat input rate of 8.8 MW_T and the corresponding energy consumption data for this boiler size are shown. Energy consumption data for the other standard boiler capacities (22, 44, 58.6, and 117.2 MW_T heat input) can be calculated by multiplying the "% increase" value shown by the desired boiler heat input rate. The results of these calculations are shown in Figure 5.1-2.

For all of the cases examined, the estimated grain loading in the low-Btu fuel gas is 0.01 g/m^3 (4 grains/10^3 scf) (Ref. 5-1). This is equivalent to around 2 ng/J (0.004 lb/10^6 Btu) in the combustion flue gases, which is below the target particulate emission control level of 13 ng/J (0.03 lb/10^6 Btu). NO_x emissions from combustion of low-Btu gas should be comparable to those arising from combustion of natural gas - 50 to 100 ng NO_x (as NO_2)/J (0.12-0.23 lb/10^6 Btu). These emissions might be reduced to even lower levels if combustion modifications are incorporated into the design of the boiler. NO_x emissions of 50 to 100 ng NO_x/J represent a 55 to 87 percent reduction in the emissions from uncontrolled coal combustion (230-400 ng/J). These levels are also in-line with the target NO_x emission

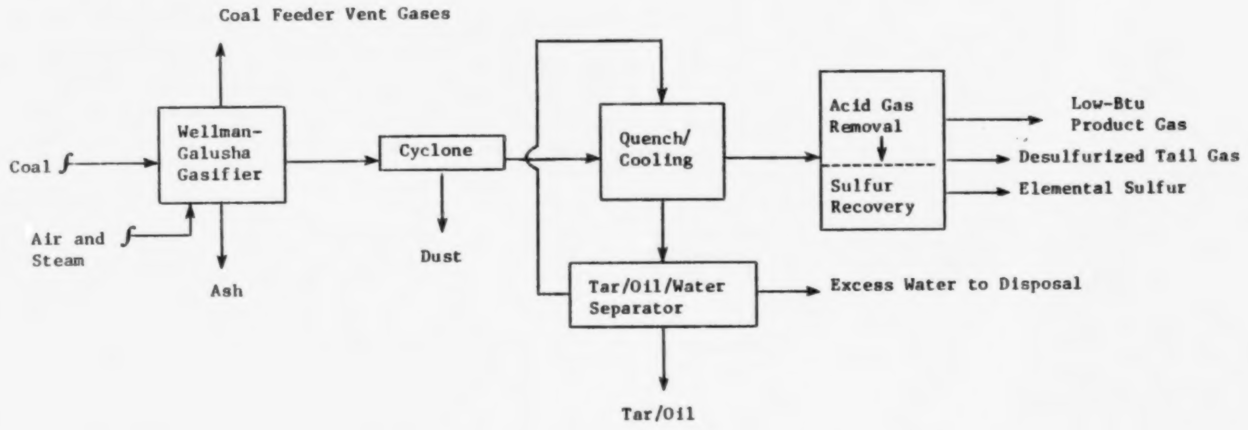

Coal Feeder Vent Gases
Coal
Air and Steam
Wellman-Galusha Gasifier
Ash
Cyclone
Dust
Quench/Cooling
Tar/Oil/Water Separator
Tar/Oil
Excess Water to Disposal
Acid Gas Removal
Sulfur Recovery
Low-Btu Product Gas
Desulfurized Tail Gas
Elemental Sulfur

Figure 5.1-1. Simplified flow diagram - Wellman-Galusha low-Btu gasification system.

TABLE 5.1-1. ENERGY CONSUMPTION FOR WELLMAN-GALUSHA LOW-BTU GASIFICATION SYSTEMS

Input Heat Rate to Boiler, MW_T (10^6 Btu/hr)*	Coal Feed	Acid Gas Removal Process and SO_2 Control Level	SO_2 Emission Control Efficiency, %	Types of Energy Consumed	Amount of Energy Consumed MW_T (10^6 Btu/hr)	% Increase in Energy Input to Uncontrolled Coal-Fired Boiler
8.8 (30)	Low Sulfur	Stretford Stringent	94.2	Coal, Steam Electricity	3.45 (11.8)	39.2
8.8 (30)	High Sulfur	Stretford Moderate	94.2	Coal, Steam Electricity	4.18 (14.3)	47.5
8.8 (30)	High Sulfur	MEA** Moderate	94.2	Coal, Electricity, Steam Low-Btu Gas	5.59 (19.1)	63.5
		Intermediate	96.5		5.59 (19.1)	63.5
		Stringent	98.2		5.60 (19.1)	63.6

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*For other boiler heat rates, energy consumed = heat rate x % increase value x 0.01

**Includes Claus to treat acid gas stream and SCOT to treat Claus tail gas.

Stringent SO_2 Control Level - 43 ng SO_2/J (0.1 lb/10^6 Btu)
Intermediate SO_2 Control Level - 86 ng SO_2/J (0.2 lb/10^6 Btu)
Moderate SO_2 Control Level - 150 ng SO_2/J (0.35 lb/10^6 Btu)

All cases have estimated NO_x and particulate emissions of:
NO_x - 50-100 ng/J (0.12-0.23 lb/10^6 Btu)
Particulates - <4 ng/J (<0.01 lb/10^6 Btu)

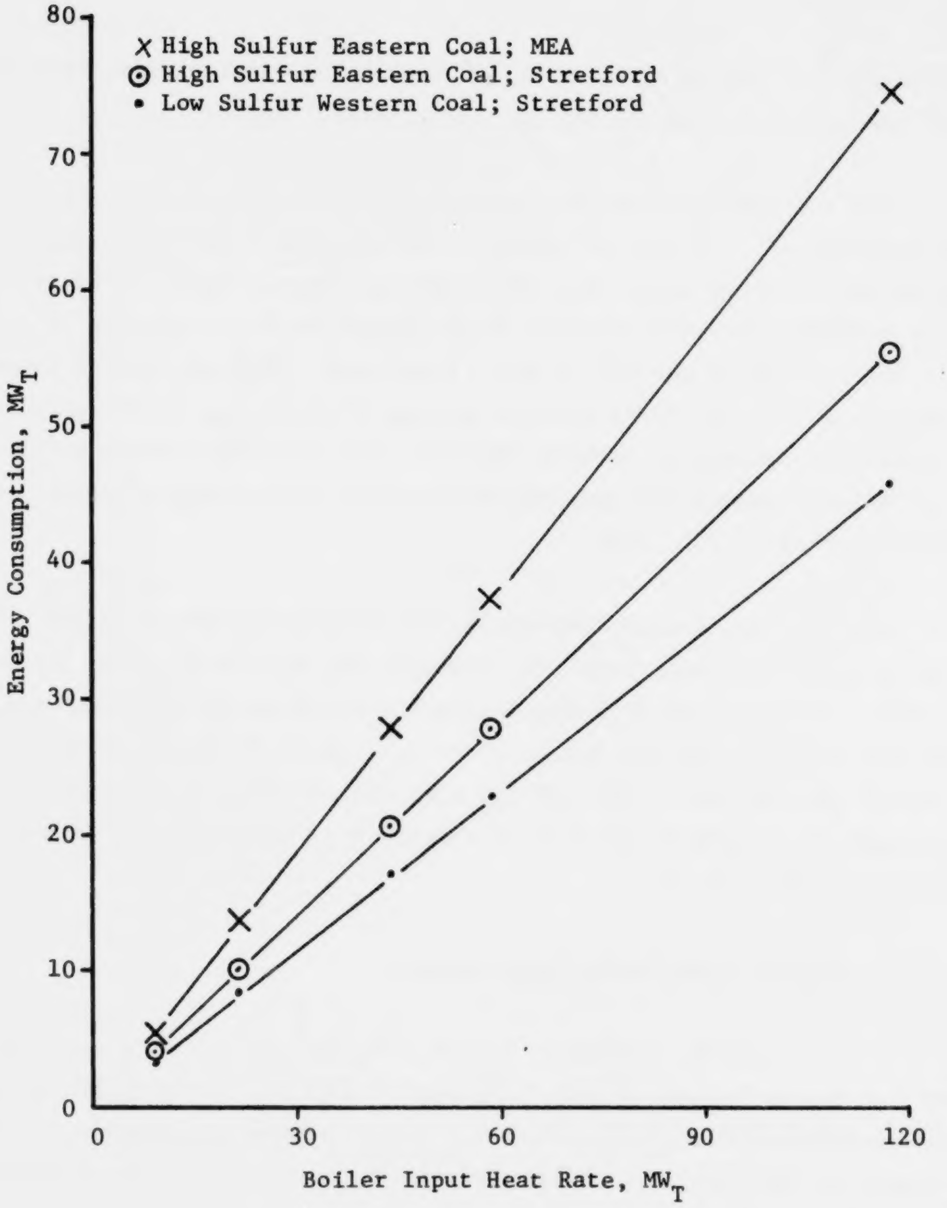


Figure 5.1-2. Energy consumption of low-Btu gasification systems.

control level (86 ng/J) (0.02 lb/10^6 Btu) being considered in this study for the combustion of low-Btu gas in industrial boilers.

For the gasification of low sulfur coal, energy requirements were calculated only for one SO_2 emission control level (corresponding to SO_2 emissions slightly lower than the stringent control level of 43 ng SO_2/J). This emission level is achieved by using the Stretford process to remove all but 10 ppmv of the H_2S in the product gas. Because the H_2S removal reaction in the Stretford process is very fast, it may not be possible to operate the process to achieve significantly less H_2S removal (although H_2S residuals of around 100 ppmv may be possible with a less efficient type of contactor) (Ref. 5-2, 5-3).

For the high sulfur coal cases, the Wellman-Galusha gasifier with the MEA acid gas removal process was examined for all three target SO_2 control levels. However, the Wellman-Galusha/Stretford system was only examined for the moderate control level. This is because of the high organic sulfur content (predominantly COS) of the high sulfur coal-derived low-Btu gas. Although the Stretford process is effective in removing H_2S, it does not remove COS (Ref. 5-4).

5.1.1 Comparison of Energy Requirements

A more detailed breakdown of the gasification system energy consumption data is shown in Table 5.1-2. A comparison of this information shows that the inefficiency of the gasification system--conversion losses and energy content of the by-product tars and oils--is the major energy consumption in all of the systems. For the low and high sulfur coal cases using the Stretford process, this represents almost 90 percent of the total energy consumed. For the higher sulfur coal cases using the MEA process, conversion inefficiency is still the major source (∿65%) of energy consumption. However, other major energy users are 1) the compressor used to compress

TABLE 5.1-2. DETAILED SUMMARY OF ENERGY CONSUMPTION FOR WELLMAN-GALUSHA LOW-BTU GASIFICATION SYSTEMS

		Low Sulfur Coal		High Sulfur Coal							
		Stretford		Stretford (Moderate Control)		MEA (Moderate Control)		MEA (Stringent Control)		MEA (Intermediate Control)	
Source of Energy Consumption	Type of Energy	kW_T	% of Product Gas	kW_T	% of Product Gas	kW_T	% of Product Gas	kW_T	% of Product Gas	kW_T	% of Product Gas
Conversion Losses	Coal Feed	1020	11.6	1970	22.4	1970	22.4	1970	22.4	1970	22.4
By-Products	Coal Feed	1960	22.3	1690	19.2	1690	19.2	1690	19.2	1690	19.2
Gasifier - Inlet Air Blower and Jacket Water Pump	Electricity*	110	1.2	100	1.1	100	1.1	100	1.1	100	1.1
Quench/Cooling											
Fan	Electricity*	130	1.5	130	1.5	130	1.5	130	1.5	130	1.5
Pumps	Electricity*	60	0.7	70	0.8	70	0.8	70	0.8	70	0.8
Electrostatic Precipitator	Electricity*	20	0.2	20	0.2	20	0.2	20	0.2	20	0.2
Condensate Evaporator	Steam**	120	1.4	30	0.3	200	2.3	200	2.3	200	2.3
Acid Gas Removal											
Gas Compressor	Electricity*					1080	12.3	1080	12.3	1080	12.3
Steam	Steam**					430	4.9	450	5.1	440	5.0
Electricity	Electricity*	30	0.3	170	2.0	<5	<0.1	<5	<0.1	<5	<0.1
Sulfur Recovery (Claus)											
Electricity	Electricity*					10	0.1	10	0.1	10	0.1
Steam	Steam**					-180	-2.0	-190	-2.1	-190	-2.1
Claus Tail Gas Treater (SCOT)											
Electricity	Electricity*					10	0.1	10	0.1	10	0.1
Steam	Steam**					40	0.4	40	0.4	40	0.4
Fuel Gas	Fossil Fuel					20	0.2	20	0.2	20	0.2
TOTAL		3450	39.2	4180	47.5	5590	63.5	5600	63.6	5590	63.5

Basis: 8.8 MW_T (30×10^6 Btu/hr) of low-Btu gas

* Electrical energy is based on energy input to the generating source and an assumed conversion efficiency to electricity of 34.1%.

**Net steam requirements are obtained from the industrial boiler.

the gas feed to the MEA unit to 0.28 MPa (40 psia) and 2) the steam used to regenerate the MEA solution.

As indicated in Table 5.1-2, the Stretford process requires much less energy than the MEA process. In fact, the quench/cooling system requires more energy than the Stretford process. Additional energy would be consumed in the Stretford process if the sulfur were melted and purified, but this would probably not be done for the small quantities of sulfur recovered in the systems considered here. Additional energy would also be required if a high temperature reduction process were used to treat the Stretford liquor blowdown, but this also is not likely for the small sizes considered in this study. Moreover, even with this increased energy consumption, the gasification inefficiency would still by the major energy consumer.

It should be noted that for the MEA cases the level of control of SO_2 emissions has a negligible impact on energy consumption. For both the moderate and intermediate control levels, energy consumption equals 63.5 percent of the low-Btu gas energy. This increases to only 63.6 percent for the stringent SO_2 control level.

5.1.2 Methods of Reducing Energy Consumption

Conversion losses and the energy content of the by-product tars and oils represent by far the greatest energy losses in converting coal into a low-Btu gaseous fuel. Thus, the greatest potential for reducing energy consumption lies with these items. Conversion losses include:

- Heat lost from the gasifier to the cooling jacket water and the atmosphere,
- Unreacted carbon in the gasifier ash,
- The sensible heat of the raw low-Btu gas, and

- The latent heat of the water vapor, tars, and oils in the raw low-Btu gas.

Little can be done about heat lost to the cooling jacket water, but fortunately this loss normally is very small.

The amount of unreacted carbon removed with the gasifier ash depends on the operation of the gasifier and the coal feed characteristics. Increasing the air flow to the gasifier helps to reduce the amount of carbon in the ash by increasing the temperature of the fire zone. However, to prevent slagging of the ash, the fire zone temperature must be kept below the ash fusion temperature. The reactivity of the coal also influences the amount of unreacted carbon in the ash. While gasifier ash from reactive bituminous coals can contain as little as 10 percent carbon (Ref. 5-1), gasifying unreactive anthracite coals can produce ashes containing over 30 percent carbon (Ref. 5-5). One means of reducing the amount of unreacted carbon in the ash is to use a slagging gasifier.

The last two sources of conversion losses listed above deal with the energy content of the gas that is lost as it is cooled prior to being treated for sulfur removal. The conventional method of recovering this type of energy is heat exchange to raise steam. However, this is normally not practical unless there is a need for low pressure steam. The use of waste heat boilers also encounters fouling problems from condensing tars and oils. A potentially more promising alternative is to not cool the gas at all, but to treat it for sulfur removal at the temperature at which it exists the gasifier. Unfortunately, acid gas removal processes capable of this are not commercially available.

The second major source of energy consumption in low-Btu gasification systems is the energy contained in the by-product tars and oils. As shown in Table 5.1-2, this equals about 20 percent of the energy content of the low-Btu product gas. While the tars and oils can be used as a fuel (Ref. 5-1), there are some practical considerations which may limit their use.

First, the quantity of tars and oil produced is small. For an 8.8 MW_T (30 x 10^6 Btu/hr) gasification facility using western coal, approximately 2000 kW_T (6.8 x 10^6 Btu/hr) of tars and oils are recovered. Site-specific factors will determine whether this amount of fuel can be economically used. Second, the tars and oils contain sulfur compounds which may restrict their use as a combustion fuel. One analysis of the tars and oils produced by gasifying a low sulfur bituminous coal indicated they contained 0.5 percent sulfur (Ref. 5-1). Again, site-specific factors must be considered. In order to illustrate the procedures used to calculate the energy impact data shown in Table 5.1-2, an example calculation is presented in Appendix D.

5.2 MEDIUM-BTU COAL GASIFICATION

The medium-Btu coal gasification systems being considered in this study as "best candidate" emission control systems for industrial boilers are the Lurgi gasifier with the following acid gas removal/sulfur recovery processes.

1) Rectisol/Stretford,
2) Stretford (for both sulfur removal and recovery).

Energy consumption data for these systems are presented in Table 5.2-1. As was true for low-Btu gasification, the energy consumed by a medium-Btu gasification system is linearly proportional to the plant's gas production rate. Therefore, in Table 5.2-1, only a standard boiler heat rate of 8.8 MW_T (30 x 10^6 Btu/hr) and the corresponding energy consumption data for this boiler size are shown. Energy consumption data for the other standard boiler heat rates, 22, 44, and 58.6 MW_T (75, 150, and 200 x 10^6 Btu/hr), or any other boiler heat rate, can be calculated by multiplying the "percent increase" value shown by the desired boiler heat rate. The results of these calculations are plotted in Figure 5.2-1.

For both the low and high sulfur coal cases, energy requirements were calculated for only one level of SO_2 emission control. For the Rectisol case, this corresponds to removal of H_2S and COS to less than 10 ppmv.

TABLE 5.2-1. ENERGY CONSUMPTION FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS

Input Heat Rate to Boiler, $MW_T(10^6$ Btu/hr)*	Coal Feed	Acid Gas Removal Process and SO_2 Control Level	SO_2 Emission Control Efficiency, %	Types of Energy Consumed	Amount of Energy Consumed, $MW_T(10^6$ Btu/hr)*	% Increase in Energy Input to Uncontrolled Coal-Fired Boiler
8.8 (30)	Low Sulfur	Stretford Stringent	93	Coal	6.07	69.0
8.8 (30)	Low Sulfur	Rectisol Stringent	99^+	Coal	6.18	70.3
8.8 (30)	High Sulfur	Stretford Intermediate	98	Coal	8.51	96.7
8.8 (30)	High Sulfur	Rectisol Stringent	99^+	Coal	8.40	95.4

* For other boiler heat rates, energy consumed = heat rate x % increase in energy use x 0.01.

Notes: Stringent SO_2 Control Level - 43 ng SO_2/J (0.1 lb/10^6 Btu)
Intermediate SO_2 Control Level - 86 ng SO_2/J (0.2 lb/10^6 Btu)

Rectisol systems include Stretford process for sulfur recovery.

All systems result in NO_x emissions estimated at 50-100 ng/J (0.12-0.23 lb/10^6 Btu) and particulate emissions of 4 ng/J (0.01 lb/10^6 Btu).

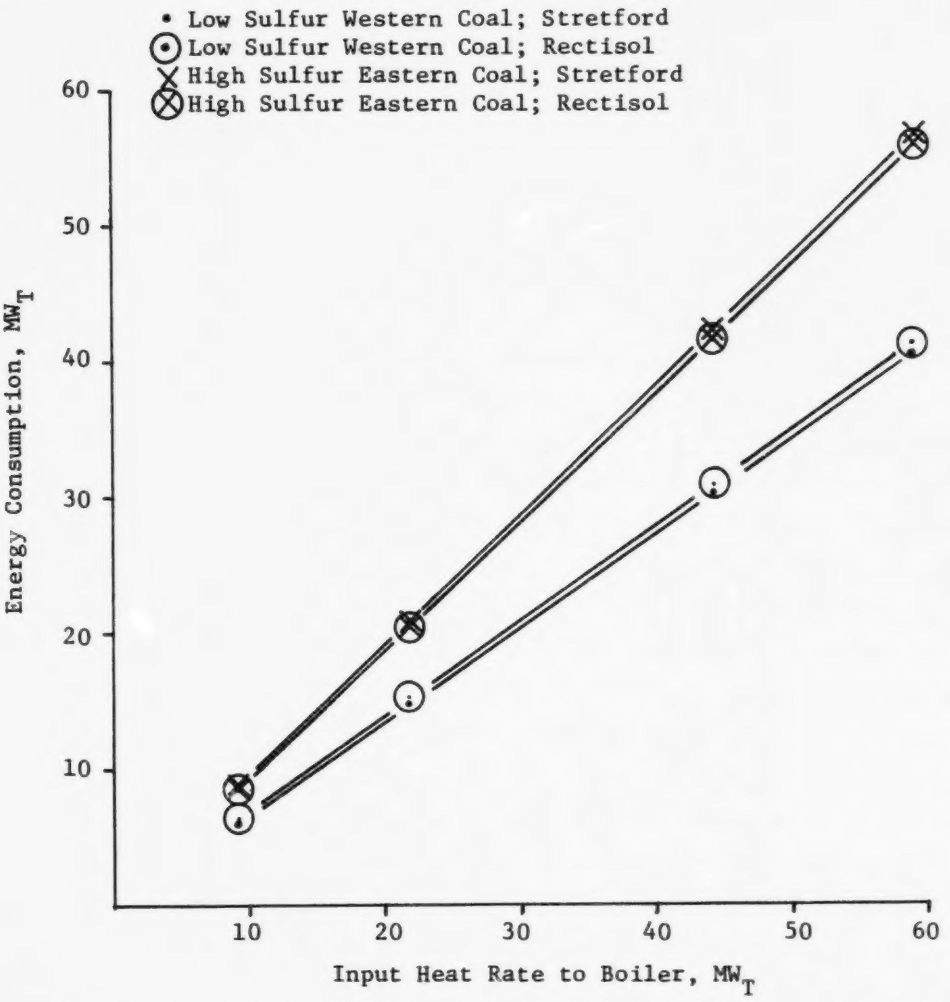


Figure 5.2-1. Energy consumption of medium-Btu gasification systems.

For the Stretford cases, COS is not removed, but H_2S is reduced to less than 10 ppmv. This was done because the medium-Btu plants examined here are assumed to be large capacity installations, 3060 MW_T (250×10^9 Btu/day), that supply gas to many users. In order to pipe medium-Btu gas and avoid corrosion problems, the H_2S concentration of the gas must be below approximately 10 ppmv (Ref. 5-6).

At total sulfur levels of less than 10 ppmv, all of the systems using the Rectisol gas removal process produce a gas that can easily meet a stringent SO_2 emission control level of 43 ng SO_2/J (0.1 lb $SO_2/10^6$ Btu). For the low sulfur coal case which uses the Stretford process for acid gas removal, approximately 160 ppmv of sulfur compounds remain in the product gas. Since this gas has a heating value of 11.6 MJ/m^3 (313 Btu/scf), its combustion will result in SO_2 emissions of less than 39 ng SO_2/J (0.09 lb $SO_2/10^6$ Btu) which is slightly below the stringent SO_2 emission control level. For the high sulfur coal case using the Stretford process, higher levels of COS increase the minimum residual sulfur level to 240 ppmv. Based on a gas heating value of 10.7 MJ/m^3 (287 Btu/scf), SO_2 emissions are estimated at 60 ng SO_2/J (0.14 lb $SO_2/10^6$ Btu). This is greater than the stringent SO_2 emission control level, but is less than the intermediate control level of 86 ng SO_2/J (0.2 lb $SO_2/10^6$ Btu).

Particulate and NO_x emissions from the combustion of medium-Btu gas should be similar to those arising from the combustion of natural gas - 4.3 ng particulates/J (0.01 lb/10^6 Btu) and 50-100 ng NO_x (as NO_2)/J (0.12-0.23 lb/10^6 Btu). These levels compare favorably with the stringent target emission levels of 13 ng particulates/J (0.03 lb/10^6 Btu) and 86 ng NO_x/J (0.2 lb/10^6 Btu).

5.2.1 Basis for Energy Consumption Data

A simplified flow diagram of a Lurgi-based medium-Btu gasification system is shown in Figure 5.2-2. Energy balances for the candidate

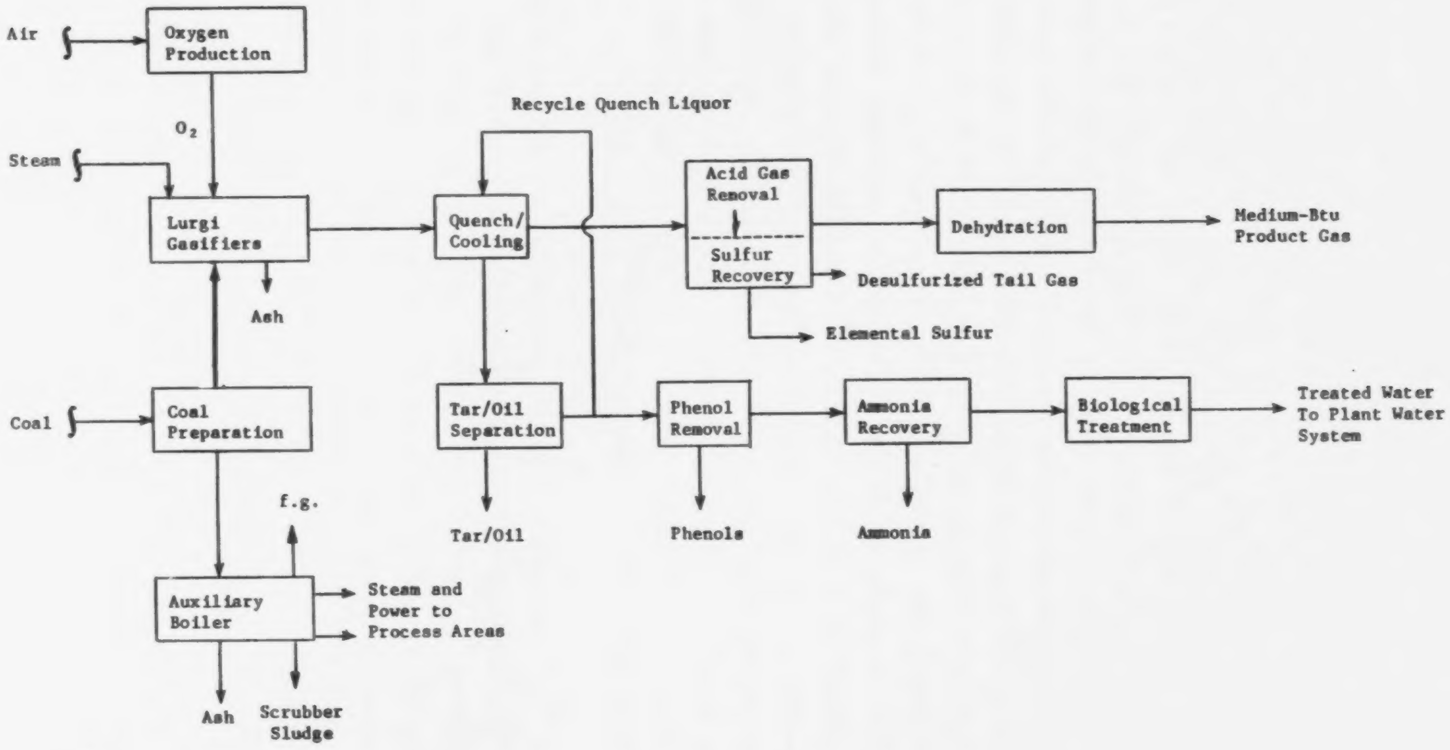

Air
Oxygen Production
O_2
Steam
Lurgi Gasifiers
Ash
Coal
Coal Preparation
f.g.
Auxiliary Boiler
Steam and Power to Process Areas
Ash
Scrubber Sludge
Recycle Quench Liquor
Quench/ Cooling
Tar/Oil Separation
Tar/Oil
Acid Gas Removal
Sulfur Recovery
Desulfurized Tail Gas
Elemental Sulfur
Dehydration
Medium-Btu Product Gas
Phenol Removal
Phenols
Ammonia Recovery
Ammonia
Biological Treatment
Treated Water To Plant Water System

Figure 5.2-2. Simplified flow scheme - Lurgi medium-Btu coal gasification system.

gasification systems using the process configuration shown in this figure were calculated from information presented in a conceptual commercial process design of a Lurgi-based high-Btu gasification plant (Ref. 5-7). In this design, all internal energy needs, i.e., steam and electricity, are supplied by an on-site coal-fired boiler.

The following adjustments were made to the high-Btu plant conceptual design in order to estimate the energy requirements of the candidate medium-Btu system:

1) The energy requirements and credits of the shift conversion, methanation and product gas compression steps were subtracted from the plant's auxiliary energy requirements.
2) The energy required by the O_2 production unit was adjusted (by linear ratioing) to reflect any difference in the O_2 requirements of the candidate systems vs the conceptual design.
3) Likewise, the auxiliary boiler load was adjusted to reflect any difference in the quantity of gasifier steam required and to reflect the absence of the methanation waste heat boiler which supplied part of the gasifier steam in the high-Btu plant design.
4) The energy required by the acid gas removal and sulfur recovery units was adjusted to reflect the use of different processes (Stretford or Rectisol).
5) A limestone flue gas desulfurization unit was used to control SO_2 emissions (90 percent control) from the coal-fired auxiliary boiler.

Estimates of the energy required by the Rectisol acid gas removal process were based on information available for a similar application in a proposed gasification plant (Ref. 5-8). In this plant, removal of 4.53 kg-moles CO_2/s and 0.0503 kg-moles H_2S/s from a stream containing 32.36 percent CO_2 and 0.34 percent H_2S had the following energy requirements:

Electricity = 8830 kW_e

High Pressure Steam = 28.39 kg/s (225,140 lb/hr)

Low Pressure Steam = 26.88 kg/s (213,180 lb/hr)

Since the composition of the inlet gas for this application is similar to the compositions of the two inlet gases being examined in this study, the above energy consumption data were divided by the quantity of acid gas removed. The resulting factors were then applied to the study cases considered here.

- 1927 kW_e per kg-mole acid gas removed per second,
- 6.194 kg high pressure steam per kg-mole acid gas removed, and
- 5.865 kg low pressure steam per kg-mole acid gas removed.

For the Stretford process, energy requirements were calculated from the following factors (Ref. 5-9).

- Electricity
  - For sulfur recovery for Rectisol cases-- 1.1 kW_e per kg sulfur removed per hour
  - For direct sulfur removal/recovery-- 3.0 kW_e per kg sulfur removed per hour
- Low Pressure Steam = 2.57 kg per kg sulfur removed per hour.

5.2.2 Comparison of Energy Consumption

A more detailed breakdown of the energy consumption data in Table 5.2-1 is presented in Table 5.2-2. For the low sulfur coal cases, the data indicate that gasification inefficiency, i.e., conversion losses and the energy content of the by-products, equals almost 35 percent of the medium-Btu product gas energy content. Stated from another viewpoint, only 70-75 percent of the energy content of the coal fed to the gasifier is converted

TABLE 5.2-2. DETAILED SUMMARY OF ENERGY CONSUMPTION FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS

	Low Sulfur Coal Feed				High Sulfur Coal Feed			
	Stretford		Rectisol/Stretford		Stretford		Rectisol/Stretford	
	MW_T	% of Product Gas	MW_T	% of Product Gas	MW_T	% of Product Gas	MW_T	% of Product Gas
Conversion Losses	540	17.6	540	17.6	630	20.5	630	20.5
By-Products	510	16.7	510	16.7	510	16.7	510	16.7
Auxiliary Energy Requirements*								
O_2 Production and Steam to Gasifiers	740	24.2	740	24.2	1350	44.1	1350	44.1
Acid Gas Removal	40	1.3	70	2.3	190	6.2	80	2.6
Sulfur Recovery			10	0.3			70	2.3
Other Process Needs	280	9.2	280	9.2	280	9.2	280	9.2
TOTAL	2110	69.0	2150	70.3	2960	96.7	2920	95.4

Basis: 3060 MW_T (250×10^9 Btu/day) of medium-Btu gas

* All steam and electricity needs not generated within the process are supplied by a 87.4% efficient coal-fired boiler.

to energy in the product gas. The other major sources of energy consumption are the oxygen production unit and gasifier steam ($\sim$24 percent of the product gas energy), and miscellaneous process needs ($\sim$9 percent). Only a small portion (1.3-2.3 percent) of the total energy needs are for the sulfur removal and recovery units which depend on the amount of H_2S removed from the gas.

The same general trends hold for the high sulfur coal cases as for the low sulfur cases. Sulfur removal and recovery account for only a small portion of the total energy consumed, while gasification inefficiency, oxygen production, and gasifier steam are the large energy users. Thus, the energy impact of utilizing medium-Btu coal gasification as an SO_2 emission control technique for industrial boilers has little dependence on the level of SO_2 emission reduction achieved.

5.2.3 Methods of Reducing Energy Requirements

The most significant potential for reducing the energy requirements of the medium-Btu gasification systems is use of the energy content of the by-product tars, oils, and naphtha. The by-products represent 15-25 percent of the total energy consumed by the system. Burning these by-products in the auxiliary boiler would reduce the coal feed to the boiler by 25-45 percent.

Another large energy requirement is steam for the gasifier. This is especially true for the high sulfur coal cases which, because of the relatively low reactivity of bituminous coals, require large amounts of steam. Use of a slagging type gasifier for the high sulfur eastern (bituminous) coal case would help reduce the gasifier steam requirements.

5.3 COAL LIQUEFACTION

Liquefaction systems employ various operations which are very similar to those used in the petroleum refining industry. Some of these include slurry mixing and preheating, high pressure hydrogenation reaction, vacuum distillation, and gas compression. Each of these operations is highly energy intensive. Pumps employed in the liquefaction process use electric energy whereas preheaters burn the process derived low-Btu fuel gas or light hydrocarbons which ultimately come from coal energy. Steam used as a heating source is also obtained by burning part of the low-Btu fuel gas or the liquid product in auxiliary boilers.

Electric power is mainly consumed by pumps, compressors, forced and induced draft fans, agitators, pressure filters, centrifuges, coal pulverizers, and lock hoppers. Table 5.3-1 shows a partial list of major electric power consuming equipment along with specific equipment functions. In most cases, all electric power will be purchased from a local utility company and the distribution system will be designed for high reliability.

Fuel gas is required for the slurry furnace, solvent hydrogenation furnace, steam reformer furnace (if steam reforming is used for hydrogen production), sulfur tail gas burner, and the off-site steam boiler. Low-Btu gas or light hydrocarbons produced within the system are used as the fuel source, although natural gas can also be used.

High, medium, and low-pressure (4.1, 2.05, and 0.9 MPa) steam is required in the liquefaction system. Steam is generally used as an indirect heating source for various liquids. In the liquefaction system extensive heat integration, where high-temperature process streams are cooled by passing boiler feed water through heat exchangers thereby generating steam, has resulted in sufficient on-site steam generation. Basically, off-site steam is required for a startup and emergency or non-normal operating mode situation. As a result a very low off-site

TABLE 5.3-1. MAJOR EQUIPMENT CONSUMING ELECTRIC POWER IN COAL LIQUEFACTION SYSTEMS

Equipment	Function
Coal-handling equipment - crushers, pulverizers	Prepared coal to the acceptable feed size
Slurry feed pump (P - 2425)	Pumps the slurry from slurry-mixing tank to the reactor via preheater
Solvent hydrogenation feed pump (P - 1675)	Pumps the solvent through fixed bed hydrotreater
Wash water booster pump (P - 1265)	Pumps the water for scrubbing the gases such as H_2S and NH_3
Air fin exchangers	Condenses the vapor in condenser and used as product coolers
Gas compressors ($P = 6.9$-15.3 MPa)	Compresses either treated gas or inert gas to the required pressures
Slurry agitator	Mixes the coal with recycled solvent in the tank
Slurry preheater-furnace draft fan	Induces the combustion gases to generate convection flow in the furnace for efficient heat transfer
Fractionation column reflux pumps	Returns the condensate from the condensate drum to the fractionation column
Fractionation column product-oil transfer pumps	Transfers the recovered oil from the fractionation towers to the downstream processing units

steam base load (boiler plant) is required. In coal liquefaction plants, off-site package boilers will be provided and will use low-Btu gas from the gasification or flexicoking operations. Coal-fired boilers are not used due to the low steam demand of the liquefaction plant and high investment requirements for coal handling equipment and flue gas desulfurization.

The overall thermal efficiency of four liquefaction systems is given in Table 5.3-2. Thermal efficiency is defined as the gross heating value of all products, divided by the gross heating value of the feed coal plus the coal equivalent of electricity. The conceptual design for SRC systems (Ref. 5-10) and the EDS system (Ref. 5-11) assume the electric power to be purchased from an outside utility. The H-Coal design (Ref. 5-12) assumes on-site power generation. The thermal efficiency is higher for the SRC-I system, primarily because of mild hydrogenation conditions in the reactor. In this case, less hydrogen is consumed in converting coal to solid fuel. In the other three liquefaction systems, mild to severe hydrogenation conditions are required. Since there is an efficiency loss in gasifying residual hydrocarbons to produce hydrogen, a high hydrogen requirement implies at least some decrease in thermal efficiency.

TABLE 5.3-2. THERMAL EFFICIENCIES OF COAL LIQUEFACTION SYSTEMS

	SRC-I	SRC-II	H-Coal (Fuel-Oil Mode)	EDS
Input				
Coal: T/D (metric)	27,270	27,270	22,730	21,800
Electric Power: MW	86	41	184	175
Output				
Pipeline Gas	-	1	-	
LPG: m^3/D	-	1,225	-	1,270
Light Oil: m^3/D	2,490	2,480	8,155	3,490
Fuel Oil: m^3/D	-	10,090	2,470	4,730
Solid Fuel: T/D (metric)	13,970	-	-	-
Thermal Efficiency: %	71.6	70.3	67.74	63.3

Source: Ref. 5-10, 5-11, 5-12

5.3.1 Discussion of Energy Consumption

The coal liquefaction systems being considered in this study as "best candidate" emission controls for industrial boilers are the SRC-I and EDS systems. Energy consumption data for those systems are summarized in Table 5.3-3 for a 58.6 MW_T (200 x 10^6 Btu/hr) industrial boiler. Energy requirements for other size boilers can be calculated by multiplying the "percentage increase" data by the desired boiler heat rate. The results of these calculations are shown in Figure 5.3-1.

A more detailed breakdown of the energy consumed in the various modules of the EDS system is presented and discussed in the following text.

The energy consumption for the EDS system is based upon estimates considering 21,818 metric tons per stream day of clean Illinois No. 6 coal being fed to a mine mouth coal conversion plant designed to produce 9,539 cubic meters per stream day of C_3^+ coal liquids. The conceptual design for this system was prepared by Exxon Research and Engineering Company (Ref. 5-11), Florham Park, New Jersey.

Electric power requirements in various modules of the process are shown in Table 5.3-4. The coal-slurry feed pump and the coal-slurry mixing tank agitator in the liquefaction module, respectively, consume 1,482 kW_e and 1,016 kW_e of electric power. The solvent hydrogenation (hydrotreating) module contains a reactor product feed pump, wash water booster pump, low-sulfur fuel oil-product pump, naphtha pump, sour water pump, and solvent-stripper bottoms pump. The total electric power consumed by these pumps is 10,940 kW_e. In the flexicoking module the gas compressors and the gasifier air blower are the major power consumers. Compressors are the main power consuming equipment in the hydrogen generation module. These compressors are used for processing the steam-reformed gas through the cryogenic units. In the gas and water treating module, solvent-charge pumps, ammonia product compressor, and transfer pumps in the phenol

TABLE 5.3-3. ENERGY CONSUMPTION FOR COAL LIQUEFACTION SYSTEMS

Heat Input to Boiler, $MW_T(10^6$ Btu/hr)*	Coal Feed	Coal Liquefaction System and SO_2 Control Level	SO_2 Emission Control Efficiency, %	Types of Energy Consumed	Amount of Energy Consumed $MW_T(10^6$ Btu/hr)*	% Increase in Energy Input to Uncontrolled Coal-Fired Boiler
58.6 (200)	High Sulfur	SRC-I Moderate	77.1	Coal and Electricity	30.8 (105)	52.2
58.6 (200)	High Sulfur	EDS Intermediate	97.0	Coal and Electricity	37.5 (128)	64.0
		Stringent**	99.0	Coal and Electricity	37.5 (128)**	64.0

* For other boiler heat rates, energy consumed = heat rate x % increase in energy use x 0.01.

** Increased sulfur control will not significantly increase energy consumption. Actual increase could not be estimated.

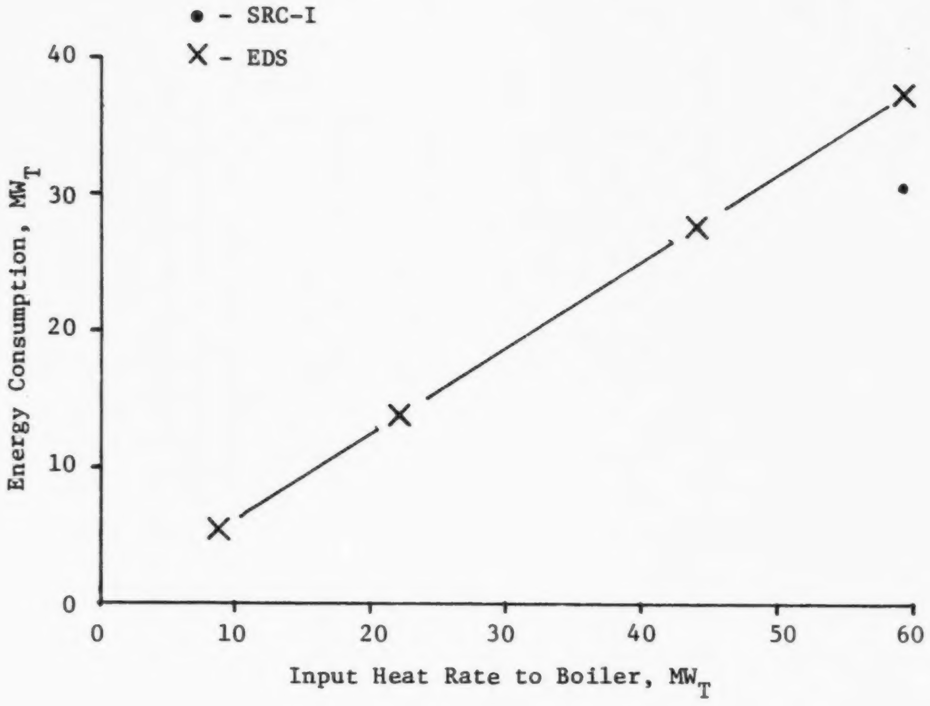


Figure 5.3-1. Energy consumption in coal liquefaction systems.

TABLE 5.3-4. ELECTRIC POWER REQUIREMENTS FOR EDS SYSTEM

Module	Normal Operating Load, kW_e
Liquefaction	24,474
Solvent Hydrogenation	11,760
Flexicoking	44,452
Hydrogen Generation (Steam reforming)	72,438
Gas and Water Treating	5,832
Product Recovery	145
Total On-Site Operating Load	159,101
Total Off-Site (Auxiliary) Operating Load	14,300
Total Plant Operating Load	173,401
	$\simeq$ 175 MW_e

extraction unit require a combined operating load of 4,942 kW_e. The pumps to transfer products such as ethane and butane that are recovered from the cryogenic units do not consume much power. The off-site or auxiliary operations include steam generation, raw water treating, boiler-feed water treating, cooling water system, wastewater treatment and reuse system, solids-waste disposal system, and coal handling and storage system. Total off-site operating load is 14,300 kW_e, which is about 9 percent of the total on-site operating load.

Table 5.3-5 shows heat equivalent of fuel gas consumption for the EDS system. Although the fuel gas requirements are met by internally generated low-Btu gas, the data in Table 5.3-5 are presented to indicate the points of energy consumption in the EDS system. Total heat consumption (from low-Btu gas with HHV of 127 Btu/scf) is about 1230 MW_T (4200×10^6 Btu/hr). If this fuel gas energy is converted to equivalent of coal energy by

TABLE 5.3-5. FUEL GAS* CONSUMPTION FOR EDS SYSTEM

Equipment	Consumption, MW_T (10^6 Btu/hr)
Slurry Furnace	410 (1400)
Solvent Hydrogenation Furnace	50 (170)
Sulfur Plant Burners	20 (70)
Steam Reformer Furnace	610 (2080)
Off-Site Boilers	140 (480)
TOTAL	1230 (4200)

* High Heating Value of Fuel Gas = 4.74 MJ/m^3 (127 Btu/scf)

assuming the coal heating value to be 29.1 MJ/kg (12,000 Btu/lb), then about 158 metric tons/hr of coal will be required to provide the necessary heat in the furnaces, sulfur burners and off-site boilers.

The conceptual design for the EDS system (Ref. 5-11) uses high pressure (4.1 MPa) and low pressure (0.9 MPa) steam in the process. Table 5.3-6 shows the steam production and consumption rates for different process modules. As for fuel gas needs, the steam requirements are supplied by internal means. This includes waste heat recovery from process streams or firing internally generated low-Btu gas in an auxiliary boiler. The numbers in parentheses in Table 5.3-6 are steam production rates. Hydrogen generation and flexicoking modules produce high pressure steam at a rate of 265,800 and 86,200 kg/hr. Atmospheric and vacuum distillation columns utilize high pressure steam at a rate of 377,000 kg/hr. The balance, 11,400 kg/hr of high pressure steam, is produced in off-site boilers. Also, as shown in the table, liquefaction and flexicoking modules produce a large amount of low pressure steam and only 55,600 kg/hr of low pressure steam is produced in the off-site boilers.

TABLE 5.3-6. STEAM REQUIREMENTS FOR EDS SYSTEM

Module	Steam Requirements,* Kg/hr 4.1 MPa Steam (600 psig)	 0.9 MPa Steam (125 psig)
Liquefaction	130,500	(22,900)
Solvent Hydrogenation	0	53,700
Flexicoking	(39,000)	(158,500)
Hydrogen Generation (Steam Reforming)	(120,400)	52,100
Gas and Water Treating	0	105,500
Product Recovery	(40,300)	300
Total On-Site Requirements	11,400	55,600

* Numbers in parentheses show steam production rates.

5.3.2 Methods of Reducing Energy Consumption

The coal liquefaction systems considered here are still in the development stage. As such, estimations of energy requirements are based on pilot and demonstration plant data and conceptual commercial size designs. Because of this, it is extremely difficult to evaluate the potential for energy savings in these systems. However, when commercialization of these systems is realized, the facility's steam and fuel gas systems will undoubtedly be optimized in order to effect the most energy efficient operation.

5.4 SUMMARY OF ENERGY IMPACTS

All of the synthetic fuels from coal systems examined in this study incur significant adverse energy impacts. Expressed as a percent of the energy input to an equivalent size, uncontrolled coal-fired industrial boiler, the energy consumption figures range from 40% to almost 100%. The low-Btu gasification systems generally have the lowest energy impacts, 40-65%, while the medium-Btu gasification systems have the highest, 70-100%. The energy impacts of the coal liquefaction systems range from 50-65%.

In all cases, the major portion of the energy consumed is attributable to conversion losses and energy contained in by-products.

References

5-1. Page, Gordon C. Environmental Assessment: Source Test and Evaluation Report--Chapman Low-Btu Gasification, Final Report. DCN 78-200-143-81. EPA-600/7-78-202. Radian Corporation, Austin, Texas. October 1978.

5-2. Telephone communication between T.G. Sipes (Radian Corp.) and John Hemmingway (Woodall-Duckham USA) concerning Stretford Process - effluent treatment and decreased H_2S removal. 24 August 1978.

5-3. Telephone communication between T.G. Sipes (Radian Corp.) and Buz Zey and Dale Williams (J.F. Pritchard and Co.) concerning cost estimates for Stretford process. 1 August 1978.

5-4. Cavanaugh, E.C., W.E. Corbett and G.C. Page. Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology, Volumes I & II, Final Report. EPA 600/7-77-125a and b. DCN 77-200-143-26, Radian Corporation, Austin, Texas. November 1977.

5-5. Thomas, W.C., K.N. Trede, and G.C. Page. Environmental Assessment: Source Test and Evaluation Report - Wellman-Galusha (Glen Gery) Low-Btu Gasification. Final Report. EPA Report No. 600/7-79-185, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. August, 1979.

5-6. Telephone communication between E.P. Hamilton (Radian Corp.) and Al Schlemmer (Texas Eastern Transmission Corp.) concerning permissible residual sulfur levels in gases entering transmission pipelines. 13 October 1978.

5-7. Detman, Roger F. Factored Estimates for Western Coal Commercial Concepts, Technical Appendix II, Engineered Equipment Descriptions, Interim Report. Alhambra, CA. C. F. Braun & Co. October 1976.

5-8. El Paso Natural Gas Co. Application of El Paso Natural Gas Co. for a Certificate of Public Convenience and Necessity. El Paso, TX. 1973.

5-9. Telephone communication between P.J. Murin (Radian Corp.) and Dale Williams and Buz Zey (J.F. Pritchard and Co.) concerning costs of Stretford process. 30 October 1978.

5-10. Schmid, B.K. and D.M. Jackson. Recycle SRC Processing for Liquid and Solid Fuels. In: Paper presented at the Fourth Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity. August 2-4, 1977. Pittsburgh, PA.

5-11. Fant, B.T. Exxon Donor Solvent Coal Liquefaction Commercial Plant Study Design. Prepared for U.S. Energy Research and Development Administration by Exxon Research and Engineering Company. Florham Park, NJ. January.

5-12. Flour Engineers and Constructors, Inc. H-Coal Commercial Evaluation - Conceptual Design and Economic Analysers for 25,000 Tons Per Day H-Coal Liquefaction Plant. Prepared for the U.S. Energy Research and Development Administration. Washington, D.C.

SECTION VI

ENVIRONMENTAL IMPACTS OF SYNTHETIC FUELS FROM COAL SYSTEMS

In previous sections of this report, the population of synthetic fuels from coal systems were identified, evaluated, and compared in order to select several "best candidate" systems for detailed analysis of cost, energy, and environmental impacts. Costs and energy impacts were addressed in Sections 4 and 5 respectively. Presented in this section are the environmental impacts associated with using synthetic fuels from coal systems as emission controls for industrial boilers.

The following text summarizes briefly the environmental impacts of synfuels processes. Detailed information is presented in Sections 6.1, 6.2, and 6.3 for the "best candidate" low-Btu gasification, medium-Btu gasification, and liquefaction systems respectively.

The synthetic fuels from coal processes being considered in this study produce both beneficial and adverse environmental impacts. Beneficial impacts include significant reductions in the emissions of SO_2 and particulates from the industrial boiler (versus those from direct coal combustion) as well as reductions in the emissions of NO_x, CO and hydrocarbons. However, the beneficial impacts realized at the boiler are offset somewhat by adverse impacts (or potential impacts) associated with producing the synthetic fuels.

For the low-Btu gasification systems, the major adverse impacts are incinerated coal feeding gases, process condensate, and gasifier ash. As coal is fed to the gasifier, raw product gas leaks past the coal feeding mechanism. Incineration of the coal feeder vent gases produces NO_x emissions as well as SO_2 emissions from the sulfur species contained in the gas. Fortunately, the flow rate of this stream is very small. A potentially more hazardous impact is the process condensate stream generated when the raw low-Btu gas is cooled. This stream contains dissolved organics, trace elements, NH_3, H_2S, HCN and other components present in the product gas. This stream in all likelihood cannot be discharged without treatment for removal of these pollutants. Ash removed from the gasifier contains trace elements (often in significant concentrations) and small amounts of organics. If this material is landfilled, it may be necessary to take steps to prevent leaching of the trace elements and organics from contaminating water sources near the landfill.

The medium-Btu gasification systems also produce coal feeding gases and solid wastes similar in nature to those from low-Btu systems. Liquid discharges are eliminated though by using evaporators for final treatment of the process condensate stream. However, the major adverse environmental impact from medium-Btu systems is the flue gas from the coal-fired auxiliary boiler. While particulates and SO_2 emissions are controlled by electrostatic precipitators (99 percent control) and limestone scrubbers (90 percent control), respectively, this stream is still the major source of air emissions from the gasification plant.

For the coal liquefaction systems, the major adverse environmental impacts are flue gases from process heaters (SO_2, NO_x, CO and particulates) and process wastewaters (oil, phenols, H_2S, NH_3 and dissolved solids).

6.1 LOW-BTU GASIFICATION

Two low-Btu gasification systems are being considered in this study as precombustion emission controls for industrial boilers. For low sulfur coal, the Wellman-Galusha gasifier with the Stretford process for sulfur removal and recovery is being examined. For high sulfur coal, the systems examined are the Wellman-Galusha gasifier with the Stretford process and the Wellman-Galusha with the MEA acid gas removal process followed by the Claus and SCOT processes for sulfur recovery.

Detailed discussions of the multimedia environmental impacts associated with the use of these systems are presented in the following sections.

6.1.1 Air Pollution

Air emissions from both the production and combustion of low-Btu gas derived from coal are summarized in Table 6.1-1. Also shown in this table are the estimated uncontrolled air emissions from direct combustion of coal in industrial boilers (Ref. 6-1). The emissions in Table 6.1-1 are expressed in terms of weight of pollutant per unit of energy input to the boiler. These data, which really are emission factors, are the same for all boiler sizes considered in this report (8.8, 22, 44, 58.6, and 117.2 MW_T heat input). This is because the actual emissions rates are linearly proportional to the boiler heat input. Actual emissions rates can be calculated by multiplying the emission factors from Table 6.1-1 by the desired boiler heat input rate.

6.1.1.1 Air Pollutants in Boiler Combustion Gases--

Particulate and NO_x emissions in the boiler combustion gases are estimated to be the same for all of the low-Btu fuel gases examined. Because low-Btu gas contains negligible amounts of particulates, particulate emissions are expected to be equal to or less than those from the combustion

TABLE 6.1-1. AIR POLLUTION IMPACTS FROM "BEST CANDIDATE" LOW-BTU GASIFICATION SYSTEMS[a]

Systems				Emissions						Other Pollutants	
				SO_2		NO_x		Particulates			
Coal Feed	SO_2 Control Level (% Reduction)	Emission Source	Sulfur Species Removal Process	ng/J	(lb/10^6Btu)	ng/J	(lb/10^6Btu)	ng/J	(lb/10^6Btu)	Pollutant	Change from Uncontrolled[b]
Low sulfur	Direct coal combustion (uncontrolled)	Combustion gas	-	520	(1.2)	340-400	(0.78-0.94)	600-1900	(1.4-4.5)		
Low sulfur	Stringent(94.2)	Combustion gas	Stretford	30	(0.07)	50-100	(0.12-0.23)	<4	(<0.01)	Organics CO Trace elements	-- - -
		Gasification system[c]		1.9	(0.005)	<0.4	(<0.001)	Small		Organics CO NH_3 HCN H_2S COS	+ + + + + +
High sulfur	Direct coal combustion (uncontrolled)	Combustion gas	-	2400	(5.7)	280-330	(0.64-0.76)	950-3100	(2.2-7.2)		
High sulfur	Moderate(94.2)	Combustion gas	Stretford	140	(0.32)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[c]		9.0	(0.02)	<0.4	(<0.001)	Small		Same as low sulfur	
	Moderate(94.2)	Combustion gas	MEA[e]	140	(0.32)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	
	Intermediate (96.5)	Combustion gas	MEA[e]	86	(0.2)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	
	Stringent(98.2)	Combustion gas	MEA[e]	43	(0.1)	50-100	(0.12-0.23)	<4	(<0.01)	Same as low sulfur	
		Gasification system[d]		13	(<0.03)	<0.4	(<0.001)	Small		Same as low sulfur	

[a] Wellman-Galusha gasifier with the indicated sulfur control processes.

[b] ++ increase; + slight increase; -- decrease; - slight decrease.

[c] Major emissions of SO_2, NO_x came from incinerated coal-feeder vent. Other emission sources are incinerated start-up gases, fugitives, and Stretford oxidizer vent.

[d] Major emissions of SO_2, NO_x came from incinerated coal-feeder vent, with smaller amounts from SCOT tail gas. Other emission sources are incinerated start-up gases and fugitives.

[e] Including Claus and SCOT for acid gas treatment.

of natural gas (4 ng/J or 0.009 lb/10^6 Btu). This represents a reduction of greater than 99 percent from uncontrolled coal-fired boiler emissions. NO_x emissions are also estimated to be similar to those from natural gas combustion (Ref. 6-2). This represents a reduction of 65-85 percent of uncontrolled coal-fired boiler NO_x emissions.

Sulfur dioxide emissions for the low sulfur coal cases are estimated at 30 ng SO_2/J (0.07 lb/10^6 Btu) based on the quantities of residual sulfur species in the synthetic gas. This is slightly below the stringent target SO_2 control level of 43 ng/J (0.1 lb/10^6 Btu). This level also represents greater than a 94 percent reduction of uncontrolled SO_2 emissions from coal-fired boilers. For the high sulfur coal cases, use of the Stretford system should result in SO_2 emission in the boiler flue gas around 140 ng SO_2/J (0.32 lb/10^6 Btu), which is less than the moderate target SO_2 control level of 150 ng/J (0.35 lb/10^6 Btu). The MEA system can be designed to achieve all three control levels. The moderate, intermediate and stringent control levels represent 94.2, 96.5 and 98.2 percent reduction, respectively, of the uncontrolled emissions from direct coal-fired boilers.

As indicated in Table 6.1-1, organic (hydrocarbon) emissions from combusting low-Btu gas are significantly lower and CO emissions are somewhat lower than would arise from direct coal combustion (Ref. 6-1). Also, emissions of trace elements may be somewhat lower. This is because many of the trace elements found in the coal feed to the gasifier appear to be concentrated in the particulates entrained in the raw low-Btu gas (Ref. 6-3). Since these particulates are removed from the gas prior to combustion, fewer should be released to the atmosphere when the low-Btu gas is combusted.

6.1.1.2 Air Pollutants from the Gasification System--

In addition to the air emissions from the combustion of low-Btu gas, various pollutants are emitted from the coal preparation, gasification, and gas purification operations of the gasification system. These emissions include:

- Incinerated coal feeding gases,
- Incinerated start-up gases,
- Stretford oxidizer vent gases,
- Tail gases from SCOT unit on MEA process,
- Fugitive emissions, and
- Evaporator emissions.

The coal feeder gas stream is mainly product gas, somewhat diluted by air, that escapes when coal is fed to the gasifier (Ref. 6-4). In most installations, this stream will probably be incinerated so that the major emissions from it will be small quantities of SO_2 and NO_x. An alternative to incineration is to recycle this stream to the gasifier inlet air or to combine it with the product gas. Since this stream accounts for most of the air emissions from the gasification system, recycling it would eliminate most of the air emissions.

Another emission source is the low-quality gas produced during start-up, shutdown, or unexpected upsets. This stream will probably be incinerated, resulting in small, intermittant emissions of SO_2 and NO_x. The quantity of these emissions cannot be estimated since they depend on the frequency of startups and upsets.

The Stretford process uses an air-blown oxidizer to convert reduced ADA back to its oxidized form. A large excess of air is used in the oxidizer and released in the oxidizer vent. The vent stream will contain primarily oxygen and nitrogen, along with water picked up from the solution. It also has the potential to contain small amounts of ammonia and hydrocarbons. If these components are present in the inlet gas, some will be absorbed by the Stretford solution and released in the oxidizer (Ref. 6-5, 6-6).

Tail gas from the SCOT unit will contain small quantities of SO_2 and H_2S, with concentrations of less than 250 ppm SO_2 and 10 ppm H_2S. Total SO_2 emissions would probably be less than 4.3 ng SO_2 per Joule of product gas (0.01 lb/10^6 Btu) (Ref. 6-7).

A vent stream from the tar/oil/water separator is a potential emission source, but is recycled to the product gas in this system and thus contributes no emissions.

Fugitive emissions from the gasification system include dust from coal handling and storage, gases discharged during ash removal and leaks around the gasifier pokeholes.

Fugitive dust from coal handling and storage consists mainly of coal dust emitted at transfer points in the handling and storage operation. This potential emission is similar to that encountered at coal-fired boilers although its magnitude would be larger due to higher coal feed rates to the gasifier. Applicable methods of control include water sprays, polymer coatings, and collection in baghouses (Ref. 6-4).

Fugitive emissions from the removal of accumulated gasifier ash are relatively small. They contain mainly steam and air, with traces of particulates and volatile matter from the ash. No special control is required.

Pokeholes are located on top of the gasifier to permit insertion of rods to monitor the depth of the ash and fire zones. This procedure occurs approximately once every four hours, with the rods being left in the pokeholes for 1-2 minutes (Ref. 6-3). During this time raw product gas leaks past the poke rods. These emissions are reduced and sometimes eliminated by decreasing the gasifier inlet air flow rate during the poking operation. This causes the gasifier to operate at slightly below atmospheric pressure, resulting in air leaking into the gasifier through the pokeholes instead of product gas leaking out. During the four hours between poking operations, the pokeholes are covered with a flat metal slide. Leaks around the pokehole cover occur. However, the magnitude of the leaks is dependent on site-specific factors such as the gasifier operating pressure and more importantly the physical condition of the slide pokehole cover, i.e., the maintenance practices of the plant.

In the low-Btu gasification systems considered in this study, excess process condensate is sent to a multiple effect evaporator for disposal. This treating step will result in the discharge of volatile components present in the process condensate. The quantities of these pollutants will depend on the amount of condensate treated and the concentration of pollutants in the condensate. No data are available on the quantities of pollutants which may be discharged from the evaporator. However, data on the composition of the process condensate are presented in the next section.

6.1.2 Water Pollution

Potential water effluents from Wellman-Galusha low-Btu gasification facilities include:

- Coal storage runoff,
- Ash sluicing water,
- Process condensate, and
- Stretford process blowdown.

Some of these potential effluents contain large quantities of specific pollutants (e.g., phenols) and cannot be discharged under the most stringent federal and state effluent standards. Controls are available, however, to reduce or eliminate the impact of these effluents.

The type and quantity of potential effluents from low-Btu gasification facilities are summarized in Table 6.1-2. Effluent rates are reported in ng/J of heat input to the boiler (lb/10^6 Btu), since effluent rates on this basis are the same for any size gasification facility. Estimated water quality parameters and compositions of the process condensate and Stretford process blowdown streams are shown in Tables 6.1-3 through 6.1-5. Coal storage runoff and ash sluicing water effluents are not characterized in Tables 6.1-2 through 6.1-5. However, these streams are essentially similar to those liquid wastes potentially discharged from coal-fired boilers,

TABLE 6.1-2. WATER POLLUTION IMPACTS FROM "BEST CANDIDATE" LOW-BTU GASIFICATION SYSTEMS[a]

Coal Feed	SO_2 Removal Level[b]	Sulfur Species Removal Process	Effluent Generated	
			Type	Quantity, ng/J ($lb/10^6$ Btu)
Low Sulfur	94.2%	Stretford	• coal pile runoff	• not quantified[c]
			• ash sluicing water	• not quantified[d]
			• process condensate	5600 (13)[e,f]
			• Stretford blowdown	300 (0.7)[f,g]
High Sulfur	94.2-98.2%	Stretford or MEA	• coal pile runoff	• not quantified[c]
			• ash sluicing water	• not quantified[d]
	94.2%	Stretford	• process condensate	1400 (3.4)[e,f]
			• Stretford blowdown	610 (1.4)[f,g]
	94.2-98.2%	MEA	• process condensate	9100 (21)[e,f]

[a] Wellman-Galusha gasification system.

[b] Effluents are the same for moderate, intermediate and stringent control levels.

[c] The quantity of water runoff from coal storage piles is largely site dependent; where coal is stored in small quantities, covered bins are usually employed, resulting in no water runoff from coal storage.

[d] The quantity of water used for ash sluicing varies widely, depending on the practices of the individual gasifier operators. The ash may be removed essentially dry, with the addition of little water. As a worst case, 1 kg of water may be used per kg of ash removed. This results in ash sluicing water of 3400-5600 ng/J (8-13 $lb/10^6$ Btu).

[e] The quantities of process condensate depends on the water content of the coal feed, the desired temperature of cooled product gas, and the pressure of the product gas. Systems featuring the MEA process produce a gas at 280 kPa (40 psia). Values shown are the quantities of condensate sent to on-site multiple effect evaporators. Residual wastes from the evaporators may be as little as 5% of the values shown.

[f] The estimated compositions of process condensate and Stretford blowdown are shown in Tables 6.1-3, 6.1-4, and 6.1-5.

[g] Based on: complete conversion of HCN in feed gas to NaCNS; 2% conversion of H_2S to $Na_2S_2O_3$; purge when salts concentration reaches 25% (Ref. 6-5, 6-8, 6-9).

TABLE 6.1-3. WATER QUALITY PARAMETERS OF QUENCH LIQUOR FROM A CHAPMAN FIXED-BED ATMOSPHERIC-PRESSURE GASIFIER[a]

Parameter	Value	Most Stringent Federal or State Effluent Standard[c]
Color (Pt-Co units)	5,000	
Odor (Threshold No.)	4,000	
pH	7.66	5-10
TDS (ppm)	6,300	
TSS (ppm)	144[b]	15 mg/ℓ
COD (ppm)	22,200[b]	125 mg/ℓ
BOD (ppm)	6,530[b]	30 mg/ℓ
DO (ppm)	0	
Conductivity (μmhos)	32,000	
Alkalinity (as $CaCO_3$)(ppm)	2,140	
Acidity	0	

[a] The process condensate from Wellman-Galusha gasification systems should have water quality similar to that shown here for a Chapman-Wilputte system.

[b] A water quality parameter is underlined once if it exceeds the most stringent federal or state effluent standard by one order of magnitude or less. A parameter underlined twice exceeds the standard by two or more magnitudes.

[c] In addition to water effluent standards, specific water quality criteria apply to receiving waters. These include:

DO – $\geq$ 5.0 mg/ℓ (probable average)
Odor Threshold Number – $\leq$ 3.0 mg/ℓ (probable average)
Alkalinity – 30-500 mg/ℓ

Source: Ref. 6-10

TABLE 6.1-4. COMPOSITION OF QUENCH LIQUOR FROM A CHAPMAN FIXED-BED ATMOSPHERIC-PRESSURE GASIFIER[a]

Constituent	Estimated Concentration (mg/ℓ)	Most Stringent Effluent Standards (mg/ℓ)
Aliphatic Hydrocarbons	30	
Glycols, Epoxides	100	
Carboxylic Acids and their derivatives	100	
Thiols	500	
Phenols	300[b]	0.005
Fused Aromatic Hydrocarbons and their derivatives	100	
Heterocyclic Nitrogen Compounds	500	
Heterocyclic Sulfur Compounds	50	
Antimony	0.07	
Arsenic	0.8[b]	0.05
Barium	0.3	1.0
Boron	9[b]	1.0
Bromine (as Br^-)	0.3	
Cadmium	0.005	0.01
Cerium	0.003	
Cesium	0.001	
Calcium	20	
Carbon		
• carbonate	2000	
Chlorine		
• chloride	3000[b]	250
Copper	0.01	0.10
Fluorine		
• Fluoride	200[b]	1
Iron	1	0.3
Iodine (as I^-)	0.3	

(Continued)

TABLE 6.1-4. Continued

Constituent	Estimated Concentration (mg/ℓ)	Most Stringent Effluent Standards (mg/ℓ)
Lanthanum	0.007	
Lithium	0.003	
Magnesium	2	
Mercury	0.0000003	0.002
Nitrogen		
• ammonia	5000[b,c]	2.5
• cyanide (alkali)	1000[b,c]	0.02
Niobium	0.008	
Phosphorous	20	1.0
Potassium	20	
Rubidium	0.01	
Scandium	0.002	
Selenium	2[b]	0.01
Silicon	2	
Silver	0.002	0.05
Strontium	0.08	
Sulfur	<8	
• thiocyanate	0.07	
• sulfide	<10	
• sulfate	1000[b]	600
Tin	0.03	
Titanium	0.1	
Tungsten	0.02	
Yitrium	0.005	
Zirconium	0.01	

[a] The process condensate from Wellman-Galusha gasification systems will have composition similar to that shown here for a Chapman-Wilputte system. The concentrations of sulfur species will be larger for systems using high-sulfur coal feeds.

[b] A concentration is underlined once if it exceeds the most stringent federal or state effluent standard by one order of magnitude or less. A concentration underlined twice exceeds the standard by two or more magnitudes.

[c] These concentrations include both free ions and complexes.

Source: Ref. 6-10

TABLE 6.1-5. COMPOSITION OF BLOWDOWN FROM STRETFORD PROCESS[a]

Constituent, mg/ℓ	Coal Feed	
	Low Sulfur	High Sulfur
$NaHCO_3$	25,000	25,000
Na_2CO_3	5,300	5,200
$NaVO_3$	4,400	6,600
ADA	6,700	10,000
Iron	50[b]	50[b]
EDTA	2,700	2,700
$Na_2S_2O_3$	45,000	120,000
NaCNS	173,000	90,000

[a] Based on: complete conversion of HCN in gas feed to NaCNS; 2% conversion of H_2S to $Na_2S_2O_3$; and salts concentration of 25%. (Ref. 6-5, 6-8, 6-9).

[b] Exceeds most stringent state or federal effluent standard.

although the quantity of coal pile runoff could be larger due to the greater coal feed requirements of gasification systems. The other streams are not found at coal-fired boilers, and thus represent incremental discharge potentially associated with low-Btu gasification facilities.

6.1.2.1 Coal Storage Runoff--

This potential effluent stream principally contains dissolved metals and inorganics that have been leached from coal in uncovered storage piles or bins. The quantity and composition of this stream are highly dependent on the site of the gasification facility, and have not been estimated for this study. A similar discharge stream exists at uncontrolled coal-fired boilers.

Runoff from coal stored in uncovered bins can be eliminated by covering the coal bins. Alternately, runoff from coal stored in uncovered bins or in piles on the ground can be collected in ditches and subsequently used in the gasification facility. Leaching of constituents from coal can be reduced by coating the coal pile with a mixture of polymers (Ref. 6-4).

6.1.2.2 Ash Sluicing Water--

Ash sluice water is used to aid the removal of ash from the gasifier. The quantity of water used for ash sluicing varies widely, depending on the practices of individual gasifier operators. As a worst case, 1 kg of water may be used per kg ash removed. If so, then the ash sluicing water would amount to about 3400 ng/J (8 lb/10^6 Btu) for the low sulfur coal cases and 5600 ng/J (13 lb/10^6 Btu) for the high sulfur coal cases.

The ash sluice water principally contains dissolved metals and inorganics that have been leached from the ash, but also contains some organic compounds. The composition of the ash sluice water depends, of course, on the characteristics of the gasifier ash. Important factors are the concentration of trace metals, unreacted coal, and the leachability of these constituents. The only data presently available on ash sluice water composition are for gasifying anthracite coal. Those data indicate few compounds are present in hazardous concentrations (Ref. 6-3). Generalizing these results to other coal types is not warranted at this time.

If plant service water is used for ash sluicing, the sluice water may be suitable for discharge into a public sewer. However, concentrations of the various trace elements and other components (such as SCN^- and CN^-) in the sluice water may be too high to allow discharge to a sewer.

If the sluice water cannot be discharged, it can be drained from the ash, collected, and recycled. The amount of sluice water that remains with the disposed ash may provide a sufficient blowdown to prevent excessive buildup of suspended or dissolved solids.

6.1.2.3 Process Condensate--

In cooling the raw low-Btu gas to the operating temperature and pressures of the sulfur removal processes (44°C for both the MEA and Stretford processes; essentially atmospheric pressure for the Stretford process, 280 kPa for the MEA process), water is condensed and subsequently removed from the gas quenching and cooling system. This condensate contains many of the constituents of the low-Btu gas, including nitrogen species (such as NH_4^+ and CN^-), particulates (which are relatively rich in trace elements), organics (including phenols, thiols, and polynuclear aromatic hydrocarbons), and dissolved gases. In tests conducted at a gasification facility with a gas quench/cooling system similar to the one being examined in this study, bioassay tests showed the process condensate to have a low degree of hazard for health effects, and a high degree of hazard for ecological effects (judged primarily from toxicity to aquatic life) (Ref. 6-10).

The size of the process condensate waste stream depends on the temperature and water content of the low-Btu gas, the temperature to which the gas is to be cooled, and the operating pressure of the sulfur removal process. The composition of the process condensate depends primarily on the composition of the raw low-Btu gas.

Tables 6.1-3 and 6.1-4 contain data describing the quality and composition of process condensate from a gasification facility using a low-sulfur coal feed. Higher concentrations of sulfur species would probably be found in the condensate from a facility using high-sulfur coal. As shown in these tables, the process condensate contains numerous constituents in concentrations exceeding the most stringent federal and state standards for effluent discharge. Those constituents include organics (such as phenols) and inorganics (such as arsenic, boron, chloride, fluoride, iron, ammonia, cyanide, phosphorous, selenium, and sulfate). The organic fraction contains a relatively large amount of polycyclic aromatic hydrocarbons, many of which are potential carcinogens.

Due to the high concentrations of organics and inorganics, this stream cannot be discharged directly to a sewer or receiving water. No similar type of potential waste stream exists at coal-fired boilers. Methods for disposal of this waste principally include:

- Evaporation,
- Containment for treatment and ultimate disposal off-site, and
- Incineration.

The selection of disposal method depends on the size and location of the gasification facility, and on the availability of waste treatment facilities near the gasification plant. For small gasifiers, the quantity of process condensate is too small to justify the construction of new on-site treatment facilities. If the industrial complex in which the gasifier is located already contains a central waste treatment plant, the process condensate could be treated there. Alternately, the condensate could be contained on-site with subsequent shipment to a large hazardous waste treatment plant located off-site. Treatment at off-site waste treatment facilities would include destruction or recovery of organics and the neutralization or recovery of inorganics. Ultimate disposal at the off-site facility would include evaporation or incineration of aqueous wastes, and ponding of waste sludges (Ref. 6-4).

Even in the larger plants considered in this study, the quantity of waste generated probably does not justify the construction of complete waste treatment facilities on-site. Again, the preferred disposal practice depends on the availability of waste treatment facilities near the gasification plant.

Evaporation and incineration control methods could be used for partial treatment of the process condensate waste stream. However, each control method has residual multimedia emissions. Evaporation effectively reduces the size of the liquid waste stream but emissions will result from the volatile components from the liquor. Non-volatile components remain with the aqueous or solid residual and require treatment and/or ultimate disposal. Incineration of the condensate waste stream, either in a process waste incinerator or in a large boiler or heater located on-site, destroys the hazardous organics contained in the waste. However, other hazardous components are volatilized and released with the incinerated gases. Residual wastes require treatment and/or ultimate disposal. In this study, excess process condensate is reduced to an estimated 5 percent of its original volume by forced evaporation.

6.1.2.4 Stretford Process Blowdown--

The Stretford process blowdown is another waste stream not found at coal-fired boilers. Its approximate compositions for the systems examined in this study are shown in Table 6.1-5.

The blowdown is required to remove nonregenerable compounds formed by the reaction of HCN to form thiocyanates and by the oxidation of hydrogen sulfide to thiosulfate. The major factors affecting the size of the blowdown stream are the HCN concentration in the feed gas and the rate of thiosulfate formation. Other factors include the degree of washing of the by-product sulfur cake and the total salts concentration in the solution. The Stretford processes operating with the gasification systems in this study produce only a partially washed sulfur cake. This reduces the blowdown rate somewhat, since some nonregenerable compounds exit the system with the sulfur cake.

The principal pollutants found in the Stretford blowdown are thiosulfate and thiocyanate. Specific standards for the discharge of these pollutants do not exist.

In the past, the liquid blowdown from the Stretford process was discharged directly to municipal sewers. Due to the high concentrations of thiocyanate and thiosulfate ions, however, the stream probably cannot now be discharged without treatment. Schemes proposed to treat Stretford blowdown include:

- Treatment and discharge,
- Regeneration, and
- Pretreatment to reduce size.

Processes involving treatment and discharge of the blowdown include biodegradation, evaporation, and combustion. These processes all require ultimate disposal of a waste stream. They also decompose the absorbing solutions, so that makeup chemicals must be added to the Stretford process.

In regeneration processes, blowdown is treated and returned to the system. The requirement for ultimate disposal of a waste stream is eliminated, and chemical makeup requirements are reduced.

The blowdown stream can be reduced in size by removing HCN from the gas before the Stretford absorber. Aqueous wastes from pretreatment contain high concentrations of thiocyanate, sorbent, ammonia, and sulfur species and require further treatment before discharge.

For small installations, the quantity of blowdown is of course very small. For small quantities of waste, the preferred treatment option is to send the blowdown to existing wastewater treatment facilities, if they are available at the gasification site. Alternately, the blowdown can be contained and subsequently shipped to large hazardous waste treatment plants located off-site. The blowdown could also be disposed of with the gasifier ash or sulfur cake; however, the subsequent potential for water pollution from the disposal of ash or sulfur may be unacceptable. For large installations, regeneration of the blowdown at high temperature may be the most desirable treatment method.

6.1.3 Solid Waste

Solid wastes generated by the Wellman-Galusha gasification systems include gasifier ash, cyclone dust, and sulfur cake from the Stretford process and pure sulfur and MEA blowdown from the MEA process. Production rates of these wastes are reported in Table 6.1-6. The production rates are reported in ng/J (lb/10^6 Btu), since the rates on this basis are the same for any size gasification facility. As shown in this table, the amount of gasifier ash generated increases with increasing coal ash content, while sulfur production increases with increasing coal sulfur content and with increased sulfur removal efficiency.

Solid waste production is considerably higher for the gasification and purification system than for an uncontrolled coal-fired boiler. The quantity of gasifier ash produced is about 100 to 700 percent greater than the bottom ash from a coal-fired boiler, mainly because of the higher coal throughput required for the gasification system and also because some of the

TABLE 6.1-6. SOLID WASTE IMPACTS FROM "BEST CANDIDATE" LOW-BTU GASIFICATION SYSTEMS[a]

Coal Feed	SO_2 Removal Level (Control Efficiency, %)	Acid Gas Removal Process	Solid Waste Generated		
			Type	ng/J	(lb/10^6 Btu)
Low sulfur	Direct Coal Combustion (Uncontrolled)	-	Bottom ash	473-1810	(1.1-4.2)
Low sulfur	1[b] (94.2)	Stretford	Gasifier ash	3760	(8.73)
			Cyclone dust	420	(.977)
			Sulfur cake	508	(1.18)
High sulfur	Direct Coal Combustion (Uncontrolled)	-	Bottom ash	775-2880	(1.8-6.7)
High sulfur	Moderate (94.2)	MEA[c]	Gasifier ash	6540	(15.2)
			Cyclone dust	367	(.853)
			Sulfur	1650	(3.83)
			MEA blowdown	116	(0.27)
	Intermediate (96.5)	MEA[c]	Sulfur[d]	1680	(3.90)
			MEA blowdown	116	(0.27)
	Stringent (98.2)	MEA[c]	Sulfur[d]	1690	(3.93)
			MEA blowdown	116	(0.27)
	Moderate (94.2)	Stretford	Sulfur cake[d]	3300	(7.67)

---

[a] Wellman-Galusha gasification system.
[b] The same system is used for moderate, intermediate, or stringent.
[c] Including Claus to treat acid gas; SCOT to treat tail gas.
[d] Ash and dust are the same for all 3 control levels.

coal ash evolves as fly ash during combustion, while most of it appears as gasifier ash in gasification. Cyclone dust and sulfur cake are additional solid waste products from the gasification system not produced from coal-fired boilers.

With the exception of the sulfur produced from the MEA/Claus system, all the solid wastes listed in Table 6.1-6 contain trace elements or other possibly hazardous compounds that could be leachable. The gasifier ash contains a variety of trace elements. Some of them, including Be, B, CO, Cr, Cu, Ge, Mn, Mo, Ni, U, and V appear to be concentrated in the ash compared to their concentrations in the coal, while other elements are volatilized in the gasification process and thus are depleted in the gasifier ash (Ref. 6-3). Tests conducted on ash from an anthracite-fired gasifier have indicated low levels of trace elements in the leachate (Ref. 6-3), but more data are required on leaching characteristics of ashes produced from gasification of bituminous coals. Because of the presence of trace elements and other possibly hazardous compounds, the gasifier ash may be classified as a hazardous toxic waste according to the Resource Conservation and Recovery Act (RCRA).

The cyclone dust is mostly carbon, but also contains various trace elements that may be leachable. Leaching tests conducted on cyclone dust from an anthracite-fired gasifier indicated fairly low levels of most trace elements in the leachate, although levels of selenium and manganese may exceed standards (Ref. 6-3). Like the ash, the cyclone dust may be classified as a hazardous toxic waste under RCRA. Because of its high carbon content, it could also be classified as a hazardous "ignitable" waste.

Sulfur recovered from the MEA tail gas by a Claus unit is relatively pure. It could possibly be marketed as a by-product, but marketing may be difficult because of the small quantity produced. If not sold, the pure sulfur could be disposed of by landfill. Sulfur produced by the Stretford process is a wet cake containing about 50 percent water, with around 4 percent total dissolved solids (Ref. 6-5). This cake contains chemicals from the Stretford solution that may be leachable from the sulfur cake. The amount of these chemicals in the cake depends on the degree and effectiveness of cake washing. This sulfur cake could be autoclaved and further purified to produce pure molten sulfur suitable for sale, but the small quantities produced in the systems considered in this study would probably make this purification economically unattractive.

The MEA blowdown consists of various compounds resulting from degradation of the solution. These include dithiocarbamates, thioureas, salts of thiosulfuric acid and formic acid, oxazolidone-2,1-(2-hydroxyethyl) imidazolidone-2 and N-(2-hydroxyethyl)¯ethylenediamine (Ref. 6-9). This blowdown will probably be classified as toxic.

The gasifier ash and sulfur cake (and possibly the cyclone dust) can be disposed of by landfill, with steps taken to prevent surface and ground water contamination from water runoff and leachate. If these wastes are declared to be toxic, practices for disposal of a toxic waste set forth in the RCRA will have to be followed. The MEA blowdown can also be landfilled following RCRA practices for disposal of a toxic waste. Methods that may be used to prevent surface and ground water contamination include the use of landfill sealants, or leachate collection systems, or interception of subsurface flow by placement of grouted slurry-trench cutoffs and/or drains upstream of the entire area of the landfill (Ref. 6-11).

The gasifier ash which has undergone fixation with lime may be made more structurally stable and resistant to leaching. Ashes with a sufficient amount of calcium and alkalinity may self-stabilize into a structurally sound, low permeability material. Ash from most bituminous coals (such as the high sulfur coal) is generally low in calcium and alkalinity and does not stabilize without lime addition. Ash from subbituminous coal generally achieves marginal self-stabilization so that lime addition may not be required for the ash from the low-sulfur coal. These generalizations, however, are derived from characteristics of ash from direct coal combustion. Gasifier ash has a higher organic content and possibly a different particle size distribution than combustion ash, and may have different stabilization characteristics.

The cyclone dust, as noted earlier, consists mostly of carbon. It can be incinerated, rather than being landfilled. In fact, under the requirements of the RCRA, landfill of the dust may not be allowed if it is classified as a hazardous "ignitable" waste. Even if landfill is allowed, burning the dust and taking advantage of its heating value would be more environmentally acceptable.

As noted earlier, the sulfur cake from the Stretford process can be autoclaved to produce a pure sulfur product. Although this may not be economically attractive if only the potential value of the sulfur is considered, it would at least produce a more environmentally acceptable waste product.

6.1.4 Other Environmental Impacts

Other potential environmental impacts of low-Btu gasification systems include noise, thermal pollution, electrical discharge, and radioactive emissions. However, no data are available on these potential impacts.

6.2 MEDIUM-BTU COAL GASIFICATION

The medium-Btu gasification systems being considered as precombustion emission controls for industrial boilers are the Lurgi gasifier with the Stretford or Rectisol acid gas removal process. Plants using these systems are assumed to be centrally located in an industrial region and are capable of producing the energy equivalent of 3060 MW_T (250 x 10^9 Btu/day) of medium-Btu gas. Industrial boiler users would receive (via pipeline) only a portion of the plant's output. Since the emissions from the gasification plant can be prorated according to the plant output, the environmental impact data presented in the following sections are expressed as weight of pollutant per unit of gas energy. Emissions associated with a given industrial boiler can be calculated from the data by multiplying by the boiler input heat rate.

6.2.1 Air Pollution

Air emissions generated during the production of medium-Btu gas are summarized in Table 6.2-1 along with estimates of the emissions resulting from combustion of the medium-Btu gas and from direct combustion of coal. NO_x and particulate emissions from the combustion of medium-Btu gas are estimated to be similar to those from combustion of natural gas - 50-100 ng NO_x/J (0.12-0.23 lb/10^6 Btu) and 4 ng particulates/J (0.01 lb/10^6 Btu) (Ref. 6-1, 6-2). This represents greater than 99 percent control of particulate emissions and 60-85 percent control of NO_x emissions (versus those from direct coal combustion). Sulfur dioxide emissions are reduced by greater than 99 percent (to 2 ng/J or 0.005 lb/10^6 Btu) by the Rectisol process for both the low and high sulfur cases. For the Stretford systems, 93 percent (37 ng/J or 0.04 lb/10^6 Btu) and 98 percent (60 ng/J or 0.14 lb/10^6 Btu) SO_2 control are achieved for the low and high sulfur coal cases, respectively.

TABLE 6.2-1. SUMMARY OF AIR POLLUTION IMPACTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS

				Emissions							
				SO_2		NO_x		Particulates		Other Pollutants	
Coal Feed	SO_2 Control Level (% Reduction)	Emission Source	Sulfur Species Removal/Recovery Process	ng/J	(lb/10^6 Btu)	ng/J	(lb/10^6 Btu)	ng/J	(lb/10^6 Btu)	Pollutant	Change from Uncontrolled*
Low sulfur	Direct coal combustion (uncontrolled)	Industrial boiler flue gas	-	520	(1.2)	340-400	(0.78-0.94)	600-1900	(1.4-4.5)		
Low sulfur	Stringent (93)	Industrial boiler flue gas	Stretford	37	(0.09)	50-100	(0.12-0.23)	4	(0.01)	Organics CO Trace elements	-- - -
		Gasification system Auxiliary boiler		17	(0.040)	137	(0.32)	6	(0.015)	Organics CO NH_3 HCN H_2S COS	+ + + + + +
Low sulfur	Stringent (99+)	Industrial boiler flue gas	Rectisol/ Stretford	2	(0.004)	50-100	(0.12-0.23)	4	(0.01)	Same as Above	
		Gasification system									
		Stretford tail gas		36	(0.083)		-		-		
		Auxiliary boiler		18	(0.043)	146	(0.34)	7	(0.016)		
High sulfur	Direct coal combustion (uncontrolled)	Industrial boiler flue gas	-	2450	(5.7)	275-330	(0.64-0.76)	950-3100	(2.2-7.2)		
High sulfur	Intermediate (98)	Industrial boiler flue gas	Stretford	60	(0.14)	50-100	(0.12-0.23)	4	(0.01)		
		Gasification system Auxiliary boiler		138	(0.32)	180	(0.42)	17	(0.040)	Same as Above	
High sulfur	Stringent (99+)	Industrial boiler flue gas	Rectisol/ Stretford	2	(0.004)	50-100	(0.12-0.23)	4	(0.01)		
		Gasification system								Same as Above	
		Stretford tail gas		60	(0.14)		-		-		
		Auxiliary boiler		142	(0.33)	194	(0.45)	18	(0.042)		

* ++ increase; + slight increase; -- decrease; - slight decrease.

The major sources of air pollution from the medium-Btu gasification systems are the auxiliary boilers, plus for the Rectisol systems, the incinerated sulfur recovery (Stretford) vent gases. The auxiliary boilers are coal-fired, and include an electrostatic precipitator for control of particulates (99 percent removal) and a limestone flue gas desulfurization unit for control of SO_2 emissions (90 percent removal). Per unit of energy input to the auxiliary boiler, the controlled boiler emissions are (Ref. 6-1).

	Low Sulfur Coal Cases	High Sulfur Coal Cases
SO_2, ng/J (lb/10^6 Btu)	52 (0.12)	245 (0.57)
Particulates, ng/J (lb/10^6 Btu)	19 (0.045)	31 (0.072)
NO_x, ng/J (lb/10^6 Btu)	405 (0.94)	327 (0.76)

The emission rates shown in Table 6.2-1 were calculated by multiplying the above data by the auxiliary boiler heat duties and dividing by the energy content of the medium-Btu product gas. As shown in Table 6.2-1, emissions from the Stretford systems are lower than those for the Rectisol systems. This is a reflection of the lower auxiliary energy requirements of the Stretford systems.

A second major source of air emissions from the systems using the Rectisol acid gas removal process and the Stretford process for sulfur recovery is the Stretford tail gas. The Stretford process can remove H_2S to below 10 ppmv; however, it is not effective in removing organic sulfur compounds such as COS (Ref. 6-9). Thus, all of the COS removed by the Rectisol process and sent to the Stretford process remains in the treated gas. This vent stream is incinerated using a clean fuel (or alternately it could be injected into the auxiliary boiler).

Other air emissions from the Lurgi medium-Btu gasification systems include:

- Hydrocarbons from the by-product storage tanks,
- Waste nitrogen from the oxygen production unit,
- Cooling tower vapors,
- Sulfur recovery (Stretford) oxidizer vent gases,
- Coal and ash hopper vent gases,
- Evaporator emissions, and
- Fugitive dust.

Table 6.2-2 summarizes the available information on the magnitude of these emissions. A brief discussion of each of these minor emissions is presented in the following text.

Emissions from the by-product storage tanks result from tank breathing, leaks, spills, and venting of tanks during filling and emptying. If control of these emissions is required to meet environmental requirements, vapor recovery systems are available which can achieve greater than 90 percent control (Ref. 6-13).

The waste nitrogen stream from the oxygen production unit will not present any environmental problems, since it is predominantly N_2 with only a small amount of CO_2.

The quantity of pollutants in the cooling tower vapors cannot be quantified. However, since treated gas quench water is used as makeup to the cooling tower, the potential exists for residual amounts of NH_3, H_2S, and hydrocarbons contained in the makeup to be stripped out in the cooling towers.

TABLE 6.2-2. SUMMARY OF MINOR EMISSIONS FROM LURGI MEDIUM-BTU GASIFICATION SYSTEMS[a]

Source	Contaminants	Contaminant Emission ng/J Product Gas ($lb/10^6$ Btu)
By-Product Tank Vents	Hydrocarbons NH_3	0.45 (0.001) 0.06 (0.0001)
Waste Nitrogen	---	---
Cooling Towers	NH_3, H_2S Hydrocarbons	trace
Stretford Oxidizer Vent	NH_3, HCN, COS, Hydrocarbons	not quantified
Lock Hopper Vents	H_2S CO Hydrocarbons	0.45 (0.001)[b] 26 (0.06) 9 (0.02)
Evaporator	H_2S, NH_3, Hydrocarbons	not quantified
Fugitive Dust	Coal Dust	4 (0.01)

[a] Emissions are for all gasification systems considered and for both coal feeds.

[b] Emissions are for a low sulfur coal feed; emissions will be higher for high sulfur coal feeds.

Source: Ref. 6-5, 6-6, 6-12

The Stretford oxidizer vent stream may contain ammonia and hydrocarbons. If these components are present in the inlet gas to the Stretford unit, some will be absorbed by the Stretford solution and subsequently be stripped out in the oxidizer (Ref. 6-5, 6-6). Again, however, the amounts of these pollutants cannot be quantified.

The Lurgi process uses lock hoppers for feeding coal into the pressured gasifier. In this study, the lock hoppers are pressurized with cooled product gas. After the coal charge is dropped into the gasifier, the hopper is depressurized to slightly above atmospheric pressure. The gas released during depressurization is collected in a lock gas holding vessel for subsequent recycle. However, when a fresh charge of coal is added to the lock hopper, it displaces the residual lock filling gases (Ref. 6-12).

An evaporator is included as a final treatment for the liquid wastes generated in the Lurgi systems considered in this study. The feed to the evaporator includes treated gas quench water which may contain residual amounts of NH_3, H_2S and hydrocarbons. While the amounts of these pollutants which may be vaporized and discharged to the atmosphere cannot be quantified, the potential does exist for their discharge.

Fugitive dust emissions arise from the coal crushing, grinding, conveying, etc., steps in the coal preparation operation. These emissions are controlled by installing water sprays with wetting agents at coal transfer points. Additional control could be achieved, if necessary for health or environmental reasons, by enclosing the dust sources and venting the area to a particulate collection device (Ref. 6-12).

6.2.2 Water Pollution

The major liquid effluents from the medium-Btu gasification systems considered here include:

- Condensate removed from the raw medium-Btu gas in the quench/cooling steps, and

- Blowdown from the plant's cooling water system.

However, in the conceptual commercial design upon which the systems in this study were based, liquid effluents were eliminated by recycle and reuse within the system and by use of evaporators for final treatment (Ref. 6-14). This feature of the conceptual design was retained in the medium-Btu systems in light of the requirements of the Water Pollution Control Act Amendments of 1972 which call for zero liquid discharge by 1985. Thus, there are no liquid effluents from the medium-Btu gasification systems.

6.2.3 Solid Wastes

Solid wastes generated by the Lurgi medium-Btu gasification systems include:

- Gasifier ash,
- Auxiliary boiler ash,
- Sludge from the boiler FGD unit,
- Sludge from the biological oxidation pond,
- Concentrated wastes from the wastewater evaporators, and
- Water treatment sludge.

Production rates of these wastes are shown in Table 6.2-3, in terms of weight of solid waste per Joule of product gas. By comparison, bottom ash from an industrial boiler is approximately 10-40 percent of these values. This is mainly because of the higher coal feed rates to the gasifier, the extra coal required by the auxiliary boiler and the fact that some of the ash content of the coal is released as fly ash during combustion while most of it is removed as gasifier ash in gasification systems.

TABLE 6.2-3. SUMMARY OF SOLID WASTES FROM LURGI MEDIUM-BTU GASIFICATION SYSTEMS

	Low Sulfur Coal Feed		High Sulfur Coal Feed	
Sulfur Removal Unit	Stretford	Rectisol	Stretford	Rectisol
Solid Waste, kg/s				
Gasifier Ash	10.4	10.4	17.0	17.0
Auxiliary Boiler Ash	2.5	2.5	6.6	6.9
FGD Sludge	3.2	3.5	25.6	26.9
Biopond Sludge	0.4	0.4	0.4	0.4
Wastewater Evaporator Residue	0.2	0.2	0.2	0.2
Water Treatment Sludge	<0.1	<0.1	<0.1	<0.1
Total, kg/s	16.7	17.0	49.8	51.4
Total, μg/J Product Gas	5.5	5.6	16.3	16.8
Total Ash Input to Direct Coal Fired Industrial Boiler, μg/J	2.4	2.4	3.8	3.8

Basis: 3060 MW_T (250 x 10^9 Btu/day) medium-Btu gas.

Source: Engineering calculations, Ref. 6-1, 6-14

All of the solid wastes contain trace elements, many of which may be present in significant concentrations (Ref. 6-15, 6-16). In addition, the gasifier ash contains approximately 5 percent unreacted coal (Ref. 6-17). If these wastes are ponded or landfilled, as is current practice with many solid wastes (Ref. 6-11), precautions may be necessary to prevent contamination of both surface and ground waters from solid waste leachate (Ref. 6-11, 6-15). Methods currently available include use of landfill sealants, leachate collection systems, or interception of subsurface flow by placement of grouted slurry-trench cutoffs and/or drains upstream of the entire landfill area (Ref. 6-11).

6.2.4 Other Environmental Impacts

Other potential environmental impacts of medium-Btu gasification systems include noise, thermal pollution, electrical discharges and radioactive emissions. However, no data are available on the levels of these potential impacts.

6.3 COAL LIQUEFACTION

The environmental impacts of coal liquefaction facilities may be generally classified into three problem areas: 1) pollutant releases; 2) physical disturbances; and 3) plant construction, operation, and decommission. These problem areas are generally common to all four liquefaction systems (SRC-I, SRC-II, H-Coal, EDS) discussed in previous sections.

6.3.1 Pollutant Releases

A generalized flow chart of pollutant releases is presented in Figure 6.3-1. The types of emissions can be divided into air or gaseous contaminants, aqueous contaminants, and solid wastes. Also, the emission could be either continuous or intermittent. Continuous emissions are defined as those released during normal on-stream operation, whereas intermittent

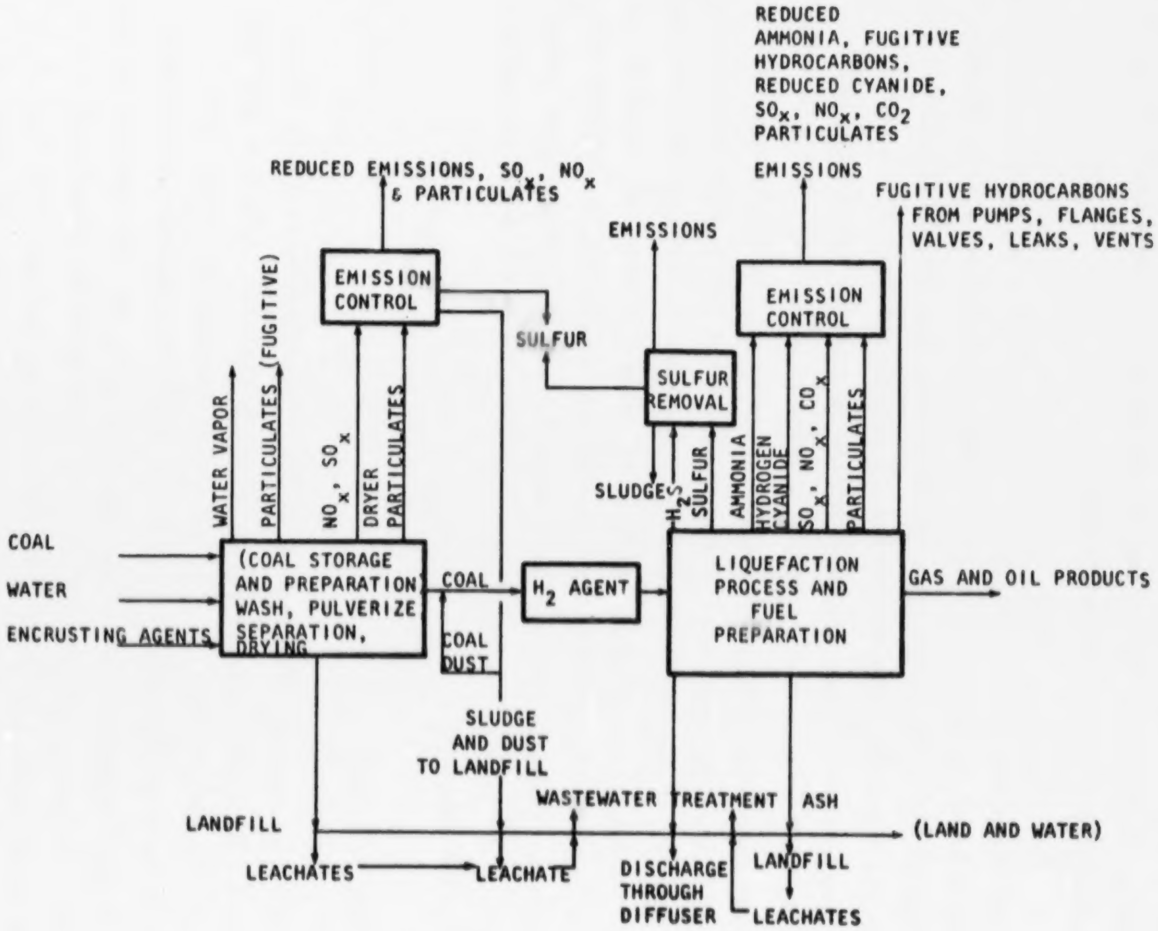

REDUCED AMMONIA, FUGITIVE HYDROCARBONS, REDUCED CYANIDE, SO_x, NO_x, CO_2 PARTICULATES
EMISSIONS
FUGITIVE HYDROCARBONS FROM PUMPS, FLANGES, VALVES, LEAKS, VENTS
REDUCED EMISSIONS, SO_x, NO_x & PARTICULATES
EMISSIONS
EMISSION CONTROL
SULFUR
SULFUR REMOVAL
EMISSION CONTROL
WATER VAPOR
PARTICULATES (FUGITIVE)
NO_x, SO_x
DRYER PARTICULATES
SLUDGE
H_2S
SULFUR
AMMONIA
HYDROGEN CYANIDE
SO_x, NO_x, CO_x
PARTICULATES
COAL
WATER
ENCRUSTING AGENTS
(COAL STORAGE AND PREPARATION) WASH, PULVERIZE SEPARATION, DRYING
COAL
H_2 AGENT
LIQUEFACTION PROCESS AND FUEL PREPARATION
GAS AND OIL PRODUCTS
COAL DUST
SLUDGE AND DUST TO LANDFILL
WASTEWATER TREATMENT
ASH
LANDFILL
(LAND AND WATER)
LEACHATES
LEACHATE
DISCHARGE THROUGH DIFFUSER
LANDFILL
LEACHATES

Figure 6.3-1. Pollutant releases from coal liquefaction system.

emissions are defined as those released during startup, shutdown, operating upsets, etc. The potential for release of hazardous chemicals, either continuously or intermittently, is one of the major environmental and safety considerations associated with coal liquefaction.

Expected rates and concentrations for air and water emissions along with expected solid waste rates for a specific liquefaction system will depend upon the size of the plant, the type of feed coal, and the final product characteristics desired. The EDS system is considered here since the conceptual environmental data for this system is readily available (Ref. 6-18), and it also represents a complex liquefaction system. The rates for air and water emissions, and solid wastes as given in the following sections, are based upon the EDS commercial plant study design for a plant producing about 9,540 m^3/SD of liquid products from 21,820 metric tons/SD of Illinois No. 6 coal. Although no attempt was made to quantify emissions release rates where insufficient process information was available, an attempt was made to at least identify all probable emission sources. Additional emission sources may be identified, however, as a result of further work and refinement of the EDS process.

6.3.1.1 Air Pollution--

Gaseous contaminants produced in a coal liquefaction facility which could have a detrimental impact on air quality include: hydrogen sulfide, ammonia, particulate matter (e.g., coal dust and process fines), hydrocarbons, sulfur dioxide, hydrogen cyanide, and small amounts of nitrogen oxides, polycyclic hydrocarbons, and heavy metals. Air emissions from the EDS process are summarized in Table 6.3-1. The greatest uncertainty with regard to those emissions is in the coal receipt, storage, and preparation area. The coal unloading, conveying, and crushing operations along with the coal storage piles are all potential sources of fugitive dust. Although

TABLE 6.3-1. ESTIMATED AIR EMISSIONS FROM THE EDS LIQUEFACTION PROCESS

	Emission Rates and Concentrations				
	SO_2[a]	CO	NO_x[b]	Hydrocarbons	Particulates
Source of Emission	kg/hr (vppm)	kg/hr (vppm)	kg/hr (vppm)	kg/hr	kg/hr (vppm)
CONTINUOUS SOURCES OF EMISSIONS					
Coal storage and handling facilities	-	-	-	-	Small but undetermined.
Slurry drier feed system	-	-	-	-	Small but undetermined.
Liquefaction slurry preheat furnaces	271 (150)	15 (20)	87 (100)	-	6 (7)
Solvent hydrogenation feed preheat furnaces	36 (150)	2.3 (20)	12 (100)		0.9 (7)
H_2 plant steam reformer furnaces	397 (150)	23 (20)	128 (100)	-	9 (7)
Off-site steam boilers[c]	41 (150)	2.3 (20)	1.3 (100)	-	0.9 (7)
Solids from flexicoker[d]	-	-	-	-	Small but undetermined.
H_2 plant deaerators	-	0.5 (6)	-	0.9	-
CO_2 removal regenerator overhead drum	-	2.7 (25)	-	-	-
Tail gas cleanup unit	18 (100)	10 (130)	1.4 (15)	-	-
Sulfur plant incinerator:					
Fuel gas combustion	1.4 (115)	(20)	0.5 (75)	-	0.5 (10)
Sulfur pit purge gas combustion	12 (1000)	0.5	-	-	-
Totals	776	56	241	0.9[e]	17
Totals, ng/J product	47	34	15	0.052	1.0
INTERMITTENT SOURCES OF EMISSIONS					
Off-site steam boilers[c]	62 (300)	1.8 (20)	50 (500)	-	-
Wastewater treating facilities - activated carbon	-	-	-	-	-
Liquefaction slurry preheat furnace - decoking	-	Small but undetermined.	-	-	10.4 (0.68 metric total)
Sulfur plant incinerator:[f]					
Fuel gas combustion	7 (45)	0.5 (20)	2.3 (30)	-	0.5 (3)
Tail gas combustion	1179 (8000)	-	5.4 (70)	-	-
Flare, safety valve discharges	g	g	g	g	g
Process vents	g	g	g	g	g

[a] The SO_2 emission rates shown in the table are based upon all sulfur being combusted to form SO_2 and no SO_3. Actually about 2-10 mol % of the sulfur can be expected to form SO_3.

[b] The NO_x emission is expected to be a 90/10 mol % mixture of NO/NO_2 as it enters the atmosphere. Most of the NO will eventually oxidize to form NO_2.

[c] The continuous emission rates shown for the off-site steam boilers are based upon the firing of low-Btu gas (LBG) as the base fuel. If insufficient LBG is available, the boilers will fire SLFO as required. The intermittent emission rates shown for the off-sites steam boilers are based upon firing LSFO as the fuel. The steam boilers have been designed to fire either fuel. If the firing of LSFO in the steam boilers proves to be unacceptable from an SO_2 and NO_x emissions standpoint, C_3 LPG can be substituted as the supplementary fuel.

[d] Flexicoker solids are sent to a settling pond which has a 5-year capacity.

[e] Hydrocarbon emissions from sources such as tankage, flaring, process and safety valve vents and equipment leakage (e.g., valves, flanges, pump seals) are not included in the total hydrocarbon emission rate. The quantity of these hydrocarbon releases and the frequency with which they might occur was not determined since this was beyond the scope of work done in the Study Design.

[f] There are two tail gas cleanup units to clean up tail gas from the three 50% sulfur plants. The emission rates shown in the table are based on only one tail gas cleanup unit being down. During this operation 50% of the total sulfur plant tail gas is sent to the other tail gas cleanup unit and the other 50% is burned in the incinerator. Some fuel gas must also be in the incinerator to support the combustion of the tail gas.

[g] Emissions can be expected from the flare, safety valves and process vents during unit startup, shutdown, or emergency situation. The frequency and quantity of these emission releases will vary and are undetermined. Some small quantity of emissions might also be expected during normal operation due to safety valve and vent safety.

the quantity of these fine particulate emissions is not known with certainty, several steps were taken in the study design to minimize these emissions. Water sprays were provided at the track hopper pit to suppress the dusting resulting from bottom-dumping of coal from railroad cars. Water sprays are also utilized at the outlets of the coal crushers. The inclined conveyor belts from the crushers to the feed distributing bins are housed in a completely enclosed gallery with emissions controlled by baghouse filters. Each transfer point along the covered conveyor belts also has a baghouse filter to remove particulates and dust.

The dry fines from the flexicoker are pneumatically conveyed to off-site mixing tanks. These dry fines are removed in a venturi scrubber prior to releasing the carrier air to the atmosphere. A very small but undetermined amount of dry fines is emitted to the atmosphere along with this air. It appears, however, that an improved processing scheme may be possible in which all the ash fines are collected in a water slurry at the flexicoker. Although not evaluated as part of the study design, this scheme may eliminate the potential dry fines emissions from the venturi scrubber while also reducing plant investment. Therefore, the emission of dry fines to the atmosphere is not likely to be a problem.

Conventional pollution control systems are provided to remove sulfur from the gas streams. An H_2S removal unit removes H_2S from the low-Btu fuel gas coming from the flexicoker. The remaining gas streams are scrubbed with DEA for H_2S and CO_2 removal. The rich DEA from on-site gas treating is fed to DEA regeneration, where H_2S and CO_2 are stripped, combined with the H_2S and CO_2 from sour water stripping, and fed to a sulfur plant. The sulfur plant converts about 95 percent of the feed sulfur to elemental sulfur. Tail gas from the sulfur plant is fed to a tail gas cleanup unit which recovers additional sulfur product. This combination of units provides for the recovery of 99.9 percent of the sulfur plant feed sulfur. The tail gas cleanup unit reduces SO_2 emissions to about 100 vppm, equivalent to new source standards for sulfur plants. There are three 50 percent sulfur plants and two 50 percent tail gas cleanup units specified in the study design. In the event that one of the tail gas cleanup units is down, 50 percent of the sulfur plant tail gas is burned in its incinerator, and the SO_2 emissions increase to about 8,000 vppm.

Continuous sulfur degasification facilities were included with the sulfur plant sulfur pits. The gases evolving from the molten sulfur, mostly H_2S, are purged with nitrogen and sent to the sulfur plant tail gas incinerator for combustion. This results in SO_2 emissions of about 1,000 vppm. However, this concentration represents an SO_2 emission rate of less than 13.6 kg/hr, which is about 1.5 percent of the total plant SO_2 emissions. The purpose of degasifying the sulfur product is to reduce the H_2S emissions during loading operations and thus reduce the risk of health hazards to personnel and odor nuisance. As an alternative to combusting the sulfur pit purge gas, it may be possible to compress it and send it to the tail gas cleanup unit for disposal. However, the scope of the work done in the study design did not permit a detailed study of this alternative. It is expected that this or some other alternative will be available as routine practice in sulfur plants in the mid-1980's.

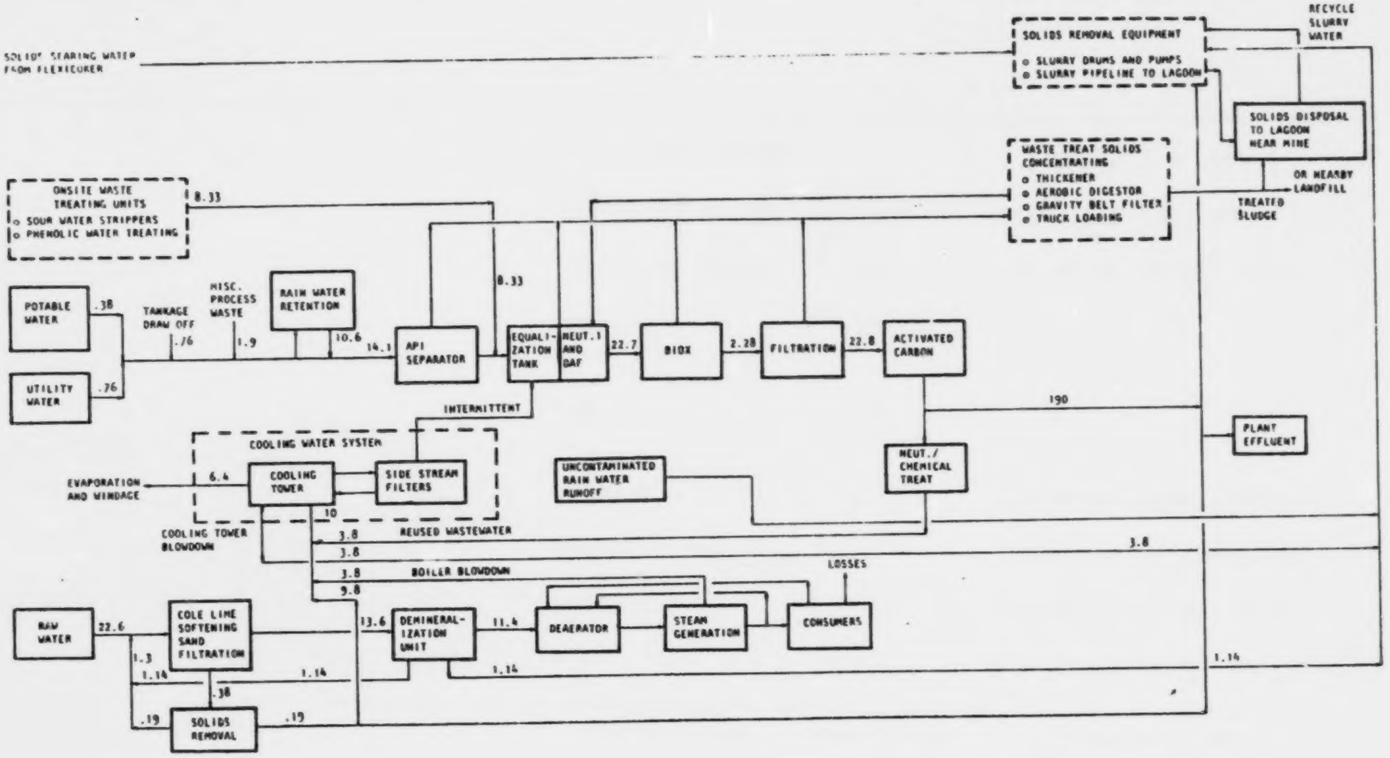
The off-site steam boilers were designed to fire low-Btu gas as a base fuel and low sulfur fuel oil (LSFO) as a supplemental fuel when required. If the firing of LSFO in these boilers proves to be unacceptable from an SO_2 and NO_x emissions standpoint, C_3 LPG can be substituted as the supplementary fuel.

6.3.1.2 Water Pollution--

Aqueous contaminants in the waste effluents which have broad temperature and pH ranges may include materials such as suspended particulates, ammonia, hydrogen sulfide, toxic trace metals, phenols, aromatic hydrocarbons, thiophenes, aromatic amines, and other organic compounds. A diagram of the wastewater treatment system for the EDS liquefaction plant is shown in Figure 6.3-2. A summary of the aqueous wastes from this plant is given in Tables 6.3-2 and 6.3-3. Also shown are the miscellaneous waste streams and applicable disposition methods.

The wastewaters from the EDS process include sour and non-sour phenolic streams. Sour water from the various EDS process blocks is combined and fed to the sour water stripping unit. The H_2S and CO_2 are stripped from the sour water and sent to the sulfur plant. Ammonia is recovered as a product in the anhydrous form. The stripped water from sour water treating is combined with the non-sour phenolic water stream coming from the slurry drier and is sent to the phenol extraction plant. This unit extracts and recovers crude phonols as a product. The effluent water along with water from the API separator is sent to the off-site wastewater equalization tank. The equalization tank serves to dampen fluctuations in wastewater composition. The API separator handles the H_2S removal unit solution purge, CO_2 removal unit solution losses, and various non-sour water streams. The total plant wastewater is drawn from the equalization tank and treated via dissolved air flotation, biological oxidation, filtration, and activated carbon. Treatment with activated carbon was included in the study design based on the assumption that the technology required to apply it will be fully available at the time it is needed for the design of a commercial EDS plant. About 15 percent of the treated wastewater is reused as cooling tower makeup, with the remainder being discharged.

The H_2S removal unit solution purge stream may present some water treating problems due to the chemical nature of this solution which contains vanadium, thiosulfates, and anthraquinone disulfonic acid (ADA). Separate treatment of this solution may be required to reduce the effluent chemical oxygen demand (COD) due to thiosulfate. Although thiosulfate is readily oxidized in a biological treatment unit, the optimum reaction occurs at low


SOLIDS SEARING WATER FROM FLEXICOKER
SOLIDS REMOVAL EQUIPMENT
o SLURRY DRUMS AND PUMPS
o SLURRY PIPELINE TO LAGOON
RECYCLE SLURRY WATER
SOLIDS DISPOSAL TO LAGOON NEAR MINE
OR NEARBY LANDFILL
TREATED SLUDGE
WASTE TREAT SOLIDS CONCENTRATING
o THICKENER
o AEROBIC DIGESTOR
o GRAVITY BELT FILTER
o TRUCK LOADING
ONSITE WASTE TREATING UNITS
o SOUR WATER STRIPPERS
o PHENOLIC WATER TREATING
8.33
8.33
POTABLE WATER
.38
UTILITY WATER
.76
TANKAGE DRAW OFF
.76
MISC. PROCESS WASTE
1.9
RAIN WATER RETENTION
10.6
14.1
API SEPARATOR
EQUALI-ZATION TANK
NEUT.1 AND DAF
22.7
BIOX
2.28
FILTRATION
22.8
ACTIVATED CARBON
190
PLANT EFFLUENT
NEUT./ CHEMICAL TREAT
UNCONTAMINATED RAIN WATER RUNOFF
INTERMITTENT
COOLING WATER SYSTEM
EVAPORATION AND WINDAGE
6.4
COOLING TOWER
SIDE STREAM FILTERS
10
COOLING TOWER BLOWDOWN
3.8
REUSED WASTEWATER
3.8
3.8
BOILER BLOWDOWN
3.8
9.8
LOSSES
RAW WATER
22.6
COLE LIME SOFTENING SAND FILTRATION
1.3
1.14
1.14
.38
.19
SOLIDS REMOVAL
.19
13.6
DEMINERAL-IZATION UNIT
11.4
DEAERATOR
STEAM GENERATION
CONSUMERS
1.14
1.14

NOTES:

1 FLOWS ARE IN m³/MIN

Figure 6.3-2. EDS commercial plant study design water systems coordination flow plan.

TABLE 6.3-2. AQUEOUS WASTES FROM OFF-SITE WASTEWATER TREATING FACILITIES OF THE COMMERCIAL EDS LIQUEFACTION PLANT

Facilities		Contaminants	Effluent Conc., ppm	Contaminant Rates
API Separator	Total plant effluent water rate = 26.5 m^3/min (average, including estimated rainwater and a capacity contingency)	Oil	5	9.1 kg/hr (0.57 ng/J)
Equalization unit		Phenols	1	1.8 kg/hr (0.096 ng/J)
Neutralization unit		Hydrogen sulfide	1	1.8 kg/hr (0.096 ng/J)
Dissolved air flotation		Ammonia	10	16 kg/hr (0.95 ng/J)
Biological oxidation		BOD	10	-
Filtration		COD	50-100	-
Activated carbon		Suspended solids	10	16 kg/hr (0.95 ng/J)
Sludge treatment		Dissolved solids	3200	121.5 T/D (303 ng/J)
		Biological sludge		7.2 T/D (18 ng/J)

TABLE 6.3-3. AQUEOUS WASTES FROM MISCELLANEOUS WASTE STREAMS OF THE COMMERCIAL EDS LIQUEFACTION PLANT

Waste Stream	Rate or Quantity	Frequency	Disposition
Spent caustic			
Sour water treating	0.015 m^3/min	Continuous	Wastewater treating following neutralization/sour water stripping/ phenol extraction
LPG prewash tower	11.3 m^3/min	Once per week	
Extraction/ regeneration	33.9 m^3/min	Once every 3 months	
H_2S removal unit solution purge	0.015 m^3/min	Continuous	Wastewater treating (API separator)
Disulfide Oil	0.001 m^3/min	Continuous	Flexicoker reactor
CO_2 removal unit solution losses	0.001 m^3/min	Continuous	Wastewater treating (API separator)
DEA system purge	378 m^3	Once per year	Scavenger/contract hauler

pH. Since the biox unit must be operated at a pH of 7 to 9 to remove organic compounds, thiosulfate may not be readily removed. If separate H_2S removal unit solution treatment is required, this may be done via acidification with sulfuric acid which converts the sodium thiosulfate to sulfate and allows for the recovery of ADA and vanadium.

6.3.1.3 Solid Wastes--

Solid wastes generated by coal liquefaction systems consist primarily of ash and refuse removed from coal and sludges, and solids recovered from waste treatment processes. Spent catalysts produced intermittently may be regenerated on-site, which could generate secondary air pollutants. Also, the disposal of solid wastes may produce fugitive dusts or leachates resulting from reaction of trace elements and ground water.

Solid wastes from the EDS process include digested biological sludge from the biological oxidation unit, oily sludge from the API separator and dissolved air flotation unit, ash from the flexicoker, and solids removed from the boiler feedwater cold lime-treating-unit blowdown. In addition, spent catalysts are disposed of on an intermittent basis. The rates of solid wastes including the catalysts to be disposed of are shown in Table 6.3-4.

Sludges from the wastewater treating facilities (API and DAF units) are thickened and then concentrated in a gravity belt filter. The sludge is then loaded onto trucks and disposed of in a land farming operation outside the plant limits. The digested biological sludge should be suitable for use in revegetating the mine tailings area and should not create any odor problems.

TABLE 6.3-4. SOLID WASTES FROM EDS SYSTEM

Continuous Source Emission	Solid Waste, Metric Tons/Day	Expected Frequency
Solids from Flexicoker	2500[a]	
Solvent hydrogenation reactors - catalyst	b	
Solvent hydrogenation feed filters gravel	135	Variable
H_2 plant hydrotreaters - catalyst	86	Once every 4 years
H_2 plant zinc oxide reactors - catalyst	216	Once every 2 years
H_2 plant steam reformers - catalyst	135	Once every 4 years
H_2 plant high temperature shift reactors - catalyst	144	Once every 2 years
H_2 plant low temperature shift reactors - catalyst	270	Once every 2 years
H_2 plant methanator reactors - catalyst	126	Once every 4 years
H_2 plant carbon treater - activated carbon	7	Once every 3 months
Sulfur plant convertors - catalyst	162	Once every 2 years
Tail gas cleanup hydrogenation reactors - catalyst	22	Once every 2 years
Alumina driers	14	Once every 2 years
Off-site steam boilers (0.2 GJ/hr)	-	When Flexicoker is down
Wastewater treating facilities - activated carbon	b	-
Liquefaction slurry preheat furnace - decoking	68	Once every 3 months
Sulfur plant incinerator:		
Fuel gas combustion	-	When one tail gas cleanup unit is down
Tail gas combustion	-	
Flare, safety valve discharges	-	Variable
Process vents	-	Variable

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[a] Flexicoker solids are sent to a settling pond which has a 5-year capacity.

[b] It is anticipated that the catalyst from solvent hydrogenation and activated carbon from the wastewater treating facilities may be returned to the manufacturer for regeneration. If either is regenerated on-site, this would represent an additional source of emissions to the atmosphere.

The ash removed from the flexicoker low-Btu product gas is slurried with water and pumped to an above-ground lagoon for disposal. This lagoon is located about one-half mile from the plant site and has about a 5-year capacity. The lagoon will eventually be covered and vegetated. The blowdown from the cold lime-treating unit is thickened and disposed of with the ash from the flexicoker. A diagram of the ash handling system is shown in Figure 6.3-3. In an actual commercial EDS plant there may be an incentive to combine ash disposal with mine tailings disposal or to dispose of the ash in the worked out portions of the mine.

The spent catalysts from the EDS process may be disposed of in several ways. These include burial or landfill, in situ or ex situ regeneration followed by reuse, or metals reclamation.

6.3.1.4 Trace Elements--

The coal fed to the EDS process contains many trace elements which may have an impact on the environment. The fate of these trace elements in the process and their potential toxicity if released to the environment need to be identified. Data are needed to give accurate and complete material balances on all trace elements. These balances will make it possible to identify potential problem areas and to design effective pollution control and disposal facilities for trace elements.

6.3.2 Physical Disturbances

The physical disturbances resulting from coal liquefaction plant construction and operation are similar to those of a coal-fired power plant. In most instances the physical disturbances are related to site-specific characteristics; however, a general inventory can be compiled and includes:

- Changes in topography due to site preparation,
- Changes in water body flow characteristics due to cooling water intake and consumptive water use,

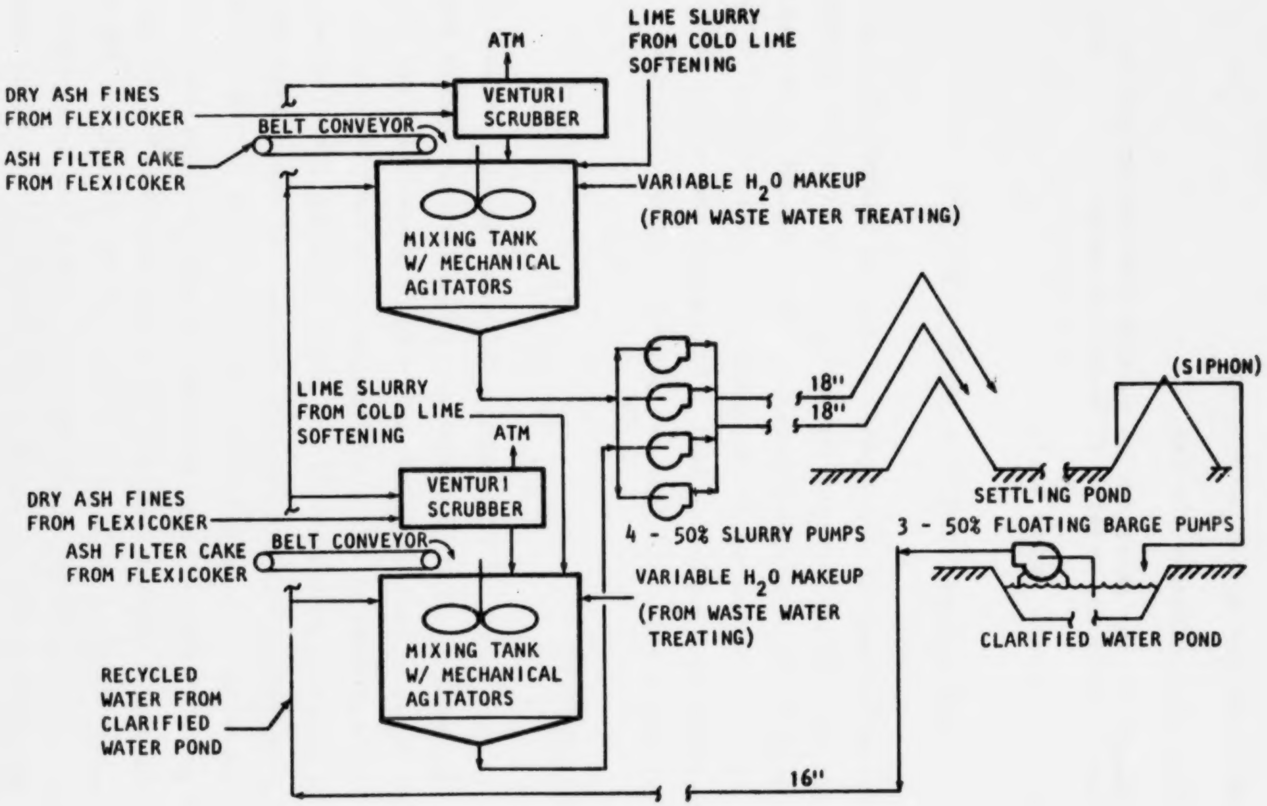
ATM
LIME SLURRY FROM COLD LIME SOFTENING
DRY ASH FINES FROM FLEXICOKER
VENTURI SCRUBBER
BELT CONVEYOR
ASH FILTER CAKE FROM FLEXICOKER
VARIABLE H_2O MAKEUP (FROM WASTE WATER TREATING)
MIXING TANK W/ MECHANICAL AGITATORS
18"
18"
(SIPHON)
LIME SLURRY FROM COLD LIME SOFTENING
ATM
SETTLING POND
DRY ASH FINES FROM FLEXICOKER
VENTURI SCRUBBER
4 - 50% SLURRY PUMPS
3 - 50% FLOATING BARGE PUMPS
ASH FILTER CAKE FROM FLEXICOKER
BELT CONVEYOR
VARIABLE H_2O MAKEUP (FROM WASTE WATER TREATING)
MIXING TANK W/ MECHANICAL AGITATORS
CLARIFIED WATER POND
RECYCLED WATER FROM CLARIFIED WATER POND
16"

Figure 6.3-3. EDS commercial plant study design ash handling system flow plan.

- Alterations of terrestrial and aquatic ecological regimes due to construction and operation,
- Aesthetic and scenic alterations,
- Minor microclimatic changes,
- Changes in the character of aquatic communities due to heated water and treated effluent discharges,
- Accumulation of solid wastes.

6.3.3 Plant Construction, Operation and Decommission

The overall effect of development of a given liquefaction system is a function not only of its physical operation but also of the socioeconomic cycle involved in construction, operation, and termination of operation of the facility. The effect would be similar to construction of a large-size power plant or a refinery plant. Throughout the entire facility life cycle, established water and land use patterns may be disrupted.

6.3.4 Other Impacts

Besides the impact on air quality and water quality created by pollutant releases as discussed earlier, other adverse impacts have been identified. Many of the compounds known to be present in coal liquefaction products are suspected of being carcinogens. Preliminary studies indicate that the products of hydrogenated and higher-boiling distillates, centrifuged oils, char, residues, and the recycled solvent oil contain potentially hazardous materials. Chronic effects of the release of low levels of these compounds and pollutant-derived atmospheric/aquatic food chain transformation products are the principal concerns for areas outside the faclities. Higher-concentration exposures and accidental discharges are the main industrial health concerns within the plant. EPA has developed a list of priority pollutants, some of which are carcinogenic. A study needs to be done to quantify the carcinogenic pollutants.

A commercial coal liquefaction facility may affect both terrestrial and aquatic indigenous plant and animal species at all levels of community organization. In most respects these impacts will be similar to coal-fired power plants.

A commercial-size coal liquefaction facility may require about 102 to 163 hectares of land. Also, disposal of solid wastes will require large areas of land. The solid wastes, depending upon the size of the plant and ash content of the coal, will vary from 1,200 to 2,800 metric tons per day. (Most of these wastes will be in the form of ash.) Over a 20-year period the solid wastes form the liquefaction plant will cover 122 to 283 hectares to a depth of about three meters. The land-use impacts and land disturbances resulting from the liquefaction facility will vary with siting requirements such as topography and access to coal supply, land area needed for the process equipment and support facilities, and secondary growth.

When coal is liquefied, hazardous compounds such as polycyclic aromatic hydrocarbons, phenols, thiophenols, aromatic amines, etc., are produced. Product streams, plant fugitive losses, and leaks may result in concentrations of toxic materials which may pose acute or chronic occupational hazards requiring specialized, strictly-controlled operating procedures and industrial hygiene and safety programs. Table 6.3-5 shows occupational safety and health hazards and control measures associated with a coal liquefaction plant. The SRC-I system is considered in the table; but similar impacts are expected from the SRC-II, H-Coal and EDS systems.

TABLE 6.3-5. OCCUPATIONAL AND HEALTH HAZARDS OF LIQUEFACTION SYSTEMS

Process Operation	Occupational Safety and Health Hazards	Nature of Hazard	Hazard Control Measures
Coal storage, preparation	Unloading	Coal dust	Water spray, filter masks
	Storage of raw coal	Water, dust, fire	Water spray, sprinklers, alarms
	Washing	Water, solids, soluble organics	Containment, water treatment
	Drying	Water, dust, organic vapors	Spray, ventilation, waste contaihment
	Crushing and pulverization	Water, dust, noise	Spray, waste containment, barriers
	Slurry preparation and pumping	Organic vapors including carcinogens	Waste containment, ventilation, contact avoidance
Liquefaction Solvent extraction Direct hydrogenation Pyrolysis Indirect liquefaction	Preheater (solvent extraction) Reactor Flash system Preheater (direct liquid) Catalytic reactor Ebullated reactor Carbonization (pyrolysis) Hydrogenation Reactor	Leakage/explosion of volatile organics and gases under heat Inhalation and skin contact w/coal tar volatiles from leaks, vents, spills Contact w/spent catalyst containing toxic or carcinogenic coal tars Maintenance and repair of vessels	Scrubbing or incineration of vent effluent Leak detection and alarm Isolation of equipment Impervious clothing and eye protection Local ventilation Air supply during vessel entry Blast protection and fire suppression
Product separation (distillation)	Centrifuge Stripper Distillation	Rotating machinery Ignition from electrical and friction sources Same as liquefaction	Safety systems, enclosure design Design Grounding
Filtration	Pressurized filter housings Lines and valves	Same as liquefaction Disposal of filtrate containing coal tar carcinogens and toxic material	Same as liquefaction Waste handling practices including containment and monitoring programs
Hydrogen production	Gasifier Steam reforming Gas separator	Explosion and fire from hydrogen leak	Leak detection Alarm system Ventilation Pressure monitors
Handling and storage of products/by-products	Pumps Storage tanks Transfer lines Valves	Leaks and explosions Inhalation/skin contact Noise	Leak detection Explosive atmosphere detection Noise monitoring Impervious clothing and eye protection Air monitoring
By-product recovery	Centrifuge Gas separator Solids removal Distillation Coking/flexicoking Decanter	Rotating machinery and ignition Leaks and explosions of pressurized gases and vapors Inhalation of toxic and carcinogenic coal tars Skin contact w/spilled residue	Leak detection Explosive atmosphere detection Noise monitoring Impervious clothing and eye protection Air monitoring
Steam/power generation	Flexicoking H_2 production	Pressure vessels Combustion products Heat Noise	Pressure monitoring Relief venting Noise monitoring Noise suppression Air monitoring
Cooling tower	Condenser, heat exchanger	Contaminated drift and blowdown Cooling tower maintenance	Leak detection Water monitoring Drift eliminator design and maintenance Equipment reliability
Water treatment	Gas separation Shift converter Decanter	Skin contact w/water containing toxic and carcinogenic coal tars Inhalation of volatile coal tar products	Water monitoring Impervious clothing and eye protection Work practices and procedures
Wastewater treatment and ponds	Centrifuge Pyrolyzer Fluidized bed combustor Flexicoking Hydrocarbonization	Rotating machinery Pressure vessels w/high temperature gases and vapors Inhalation and skin contact of coal tars	Safety systems/enclosures Landfill procedures Incineration of wastes Impervious clothing and eye protection
Solid waste disposal			
Fuel gas sulfur removal			
Gasification			
Acid gas removal			

References

6-1. Environmental Protection Agency. Compilation of Air Pollutant Emission Factors. 2nd ed. with supplements. AP-42. Research Triangle Park, NC. 1973.

6-2. Rawdon, A. H., R. A. Lisaukas and S. A. Johnson. "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames." Presented at the NO_x Control Technology Seminar, sponsored by Electric Power Research Institute, San Francisco, CA. 5-6 February 1976.

6-3. Thomas, W. C., K. N. Trede, and G. C. Page. Environmental Assessment: Source Test and Evaluation Report - Wellman-Galusha (Glen Gery) Low-Btu Gasification. Final Report. EPA Report No. 600/7-79-185, EPA Contract No. 68-02-2147. Austin, TX. Radian Corporation. August, 1979.

6-4. Cavanaugh, E. C., W. E. Corbett and G. C. Page. Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology. Volumes I & II. Final Report. EPA 600/7-77-125a, b. Radian Corporation. Austin, TX. November 1977.

6-5. Nicklin, T., and B. H. Holand. "Removal of Hydrogen Sulfide from Coke-Oven Gas by the Stretford Process." DECHEMA Monogr. 48(835-358), 243071 (1963).

6-6. Kleeberg, Ulrich. "Removal of Hydrogen Sulfide from Gases Using the Stretford Process." Presented at the Fifth International Conference on Coal Gasification, Liquefaction and Conversion to Electricity. Pittsburgh, PA. 8-10 August 1978.

6-7. Environmental Protection Agency. Standards Support and Environmental Impact Statement Volume I: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants. Research Triangle Park, NC. Emission Standards and Engineering Division. September 1976.

6-8. Moyes, A. J., and J. S. Wilkinson. "Development of the Holmes-Stretford Process." Chem. Eng. (London) 282, 84-90 (1974).

6-9. Riesenfeld, F. C., and A. C. Kohl. Gas Purification. Second edition. Gulf Publishing Company. Houston, TX. 1974.

6-10. Page, Gordon C. Environmental Assessment: Source Test and Evaluation Report--Chapman Low-Btu Gasification. Final Report. EPA-600/7-78-202. EPA Contract No. 68-02-2147. Radian Corporation. Austin, TX. October 1978.

6-11. Meserole, N. P. Review and Assessment of the Existing Data Base Regarding Flue Gas Cleaning Wastes. Radian Corporation. Austin, TX. In press.

6-12. Sinor, J. E., ed. Evaluation of Background Data Relating to New Source Performance Standards for Lurgi Gasification. Final Report. EPA-600/7-77-057. Cameron Engineers Inc. Denver, CO. June 1977.

6-13. Thomas, William C. Summary Report for a Screening Study to Determine the Emission Reduction Potential NSPS Would Have on Transfer Operations Involving Crude Oil, Jet Fuels and Aviation Gasoline. Final Report. Radian Corporation. Austin, TX. June 1976.

6-14. Detman, Roger F. Factored Estimates for Western Coal Commercial Concepts, Technical Appendix II, Engineered Equipment Descriptions. Interim Report. C. F. Braun & Company. Alhambra, CA. October 1976.

6-15. Holland, W. F., et al. Environmental Effects of Trace Elements from Ponded Ash and Scrubber Sludge. Final Report. Radian Corporation. Austin, TX. September 1975.

6-16. Radian Corporation. Coal-Fired Power Plant Trace Element Study, 4 vols. Radian Corporation. Austin, TX. September 1975.

6-17. El Paso Natural Gas Company. Application of El Paso Natural Gas Company for a Certificate of Public Convenience and Necessity. Docket No. CP73-131. El Paso, TX. 1973.

6-18. Fant, B. T. Exxon Donor Solvent Coal Liquefaction Commercial Plant Study Design. Prepared for U.S. Energy Research and Development Administration. Exxon Research and Engineering Company. Florham, NJ. January 1978.

SECTION 7

EMISSION SOURCE TEST DATA

Very little data exists on the levels of pollutants resulting from combustion of coal derived synthetic fuels in industrial boilers. Two series of tests were performed with low- and medium-Btu gas in small test furnaces. For coal-derived liquids, eight combustion tests have been conducted. The input heat rate of the coal-derived liquids test facilities varied considerably. The largest was a commercial utility boiler rated at 22.5 MW_e output or an approximate heat input of 66 MW_T (220×10^6 Btu/hr). Two tests were run on a steam boiler with an input heat rate of 14 MW_T (48×10^6 Btu/hr). The other tests were conducted on very small boilers or laboratory combustors with input heat rates of 80-800 kW_T ($0.3\text{-}3.0 \times 10^6$ Btu/hr).

The following sections give further details on the synfuels combustion tests. Included are the pollutants analyzed and the analytical methods used. Section 7.1 discusses the gaseous synfuels tests, while Section 7.2 discusses the eight coal liquids combustion tests.

7.1 LOW- AND MEDIUM-BTU GAS EMISSION TEST DATA

The Riley Stoker Corporation has conducted process development work for the Riley coal gasifier. At their test facility in Worchester, Massachusetts, they also have a small refractory lined test furnace. As part of their development and marketing of the Riley gasifier, they conducted a series of coal gas combustion tests. The results of the tests were presented (Ref. 7-1) at the EPRI sponsored NO_x Control Technology Seminar held in San Francisco, CA on February 5-6, 1976, and are summarized in Section 7.1.1.

Under a contract to the EPA, the Institute of Gas Technology has also conducted combustion tests of low- and medium-Btu gas in a small furnace. The results of those tests are summarized in Section 7.1.2.

7.1.1 Riley Morgan Combustion Tests

The internal dimensions of the test furnace used in the Riley tests were 0.91 x 0.68 x 1.6 m (36 x 27 x 62 in). The burner was a scaled register-type burner similar to those used in utility and industrial boilers. Table 7.1-1 lists the gaseous species analyzed for during the tests and the analytical methods used. As shown in this table, both NO_x and SO_2 were measured in the combustion gases. However, the major purpose of the Riley paper was to discuss NO_x emissions, and more specifically NO_x formation from fuel-bound nitrogen compounds (predominantly NH_3). Because of this emphasis, SO_2 emission data were not presented.

Most of the Riley data was for combustion of low-Btu gas (5.2-6.3 MJ/m^3) with ammonia concentrations of 280-670 ppmv. However, one test was conducted using medium-Btu gas (10 MJ/m^3) containing 1170 ppmv ammonia. The results of these tests are shown in Figures 7.1-1 and 7.1-2.

NO_x can be formed from both fuel bound nitrogen compounds and by thermal fixation of molecular nitrogen. In order to identify the quantity of NO_x attributable to thermal fixation, tests were also conducted using fuel gases with low ammonia (11-43 ppmv) concentrations. The low NH_3 concentrations were obtained by scrubbing the fuel gases with a sulfuric acid solution. The results of these tests are also shown on Figure 7.1-2. At 6 percent excess oxygen, thermal NO_x emissions were around 1.0 ppmv

TABLE 7.1-1. ANALYTICAL PARAMETERS AND ANALYSIS METHODS FOR RILEY SYNGAS COMBUSTION TESTS

Analytical Parameter	Test Method*
Product Gas Constituents	
CO	Infrared analyzer
CO_2	Infrared analyzer
CH_4	Infrared analyzer
H_2	Thermal conductivity analyzer
H_2S	Collection in NaOH followed by standard iodine titration
NH_3	Collection in H_2SO_4 followed by gas detecting ammonia electrode
Particulates and aerosols	Isokinetic sampling and collection on absolute membrane filter
Combustion Gas Constituents	
NO_x	Electrochemical cell sensors
SO_2	Electrochemical cell sensors
O_2	Electrochemical cell sensors
CO	Infrared analyzer

*All analytical parameters except H_2S and NH_3 were analyzed continuously.

Source: Ref. 7-1

Used with permission of Riley Stoker Corporation from "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames." February 1976.

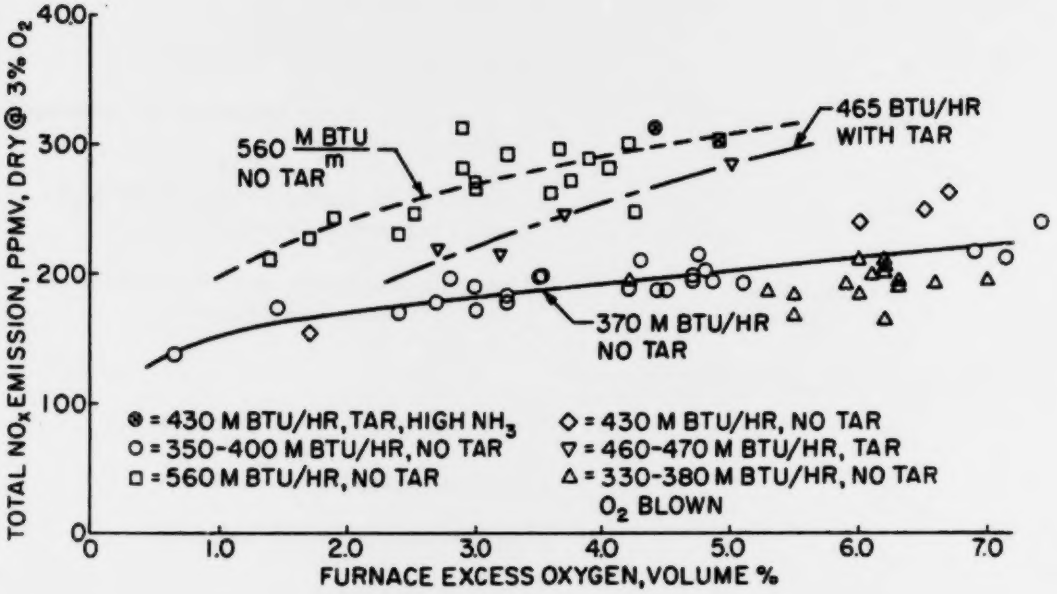

Figure 7.1-1. NO_x emission data for coal derived syngas combustion.

Source: Ref. 7-1

Used with permission of Riley Stoker Corporation from "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames." February 1976.

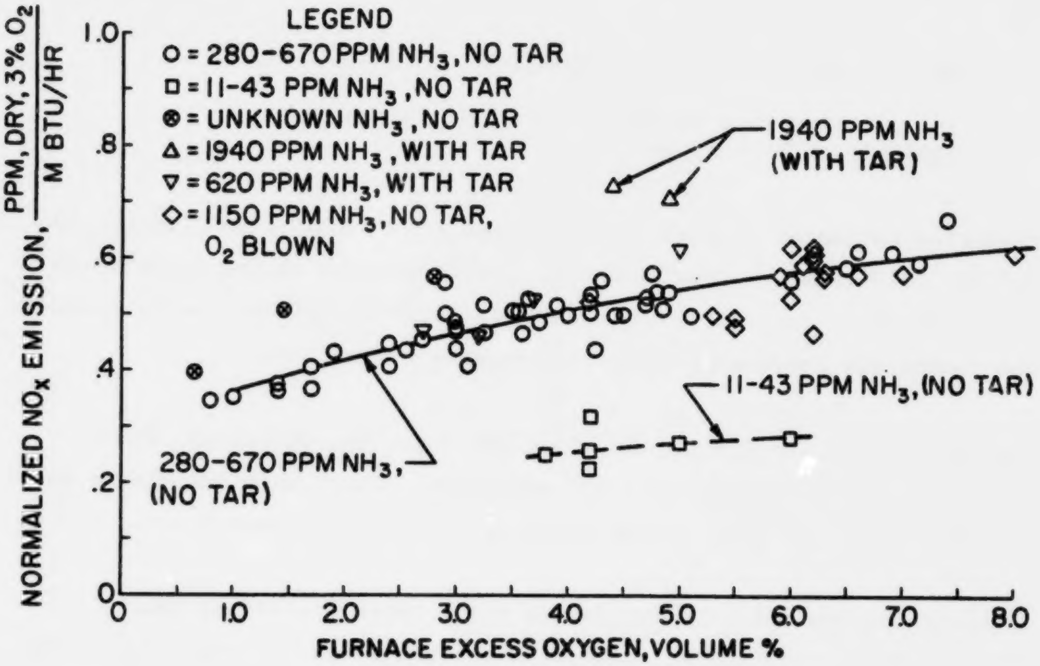

Figure 7.1-2. NO_x emission data (normalized with respect to furnace heat input) for coal derived syngas combustion.

Source: Ref. 7-1

(dry, at 3 percent O_2) per kW_T input to the test furnace. At 4 percent excess oxygen, thermal NO_x was only 0.85 ppmv (dry, at 3 percent O_2) per kW_T input. To convert these values to ng NO_x per J input to the furnace, the following factors were used:

- furnace input heat rates ranged from 101 to 173 kW_T
- flue gas flow rate is estimated at 16 g-moles per megajoule of heat input.

The resulting estimated thermal NO_x emissions are 63-120 ng NO_x (as NO_2)/J (0.15-0.28 lb/10^6 Btu). These emissions are comparable to and support the target NO_x emission control level of 86 ng/J (0.20 lb/10^6 Btu) being examined in this technology assessment report for syngas combustion.

By subtracting the thermal NO_x from the total NO_x emissions, Riley estimated the fuel-bound nitrogen NO_x emissions. These were correlated with ammonia content in the fuel gas as shown in Figures 7.1-3 and 7.1-4.

Information was not provided in the Riley paper concerning the accuracy of the reported data. Also, the rationale was not given for using electrochemical cell sensors instead of the EPA Method 7 for NO_x determinations. However, a probable reason is that using electrochemical cell sensors provided continuous data while if the EPA method was used, several hours are required to collect and analyze a single sample.

7.1.2 Institute of Gas Technology Combustion Tests

The pilot-scale furnace used by IGT was 4.3 m (14 ft) long and had a cross-sectional area of 1.98 m^2 (21.3 ft^2). Combustion tests were conducted using both a ported baffle burner and fuel momentum controller burner. The input heat rate for all tests was nominally 1 MW_T (3.5 x 10^6 Btu/hr). The low- and medium-Btu gases tested were blended to have compositions typical of gas produced from Wellman-Galusha gasifiers. Table 7.1-2 lists the analytical instruments used during the tests.

Used with permission of Riley Stoker Corporation from "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames." February 1976.

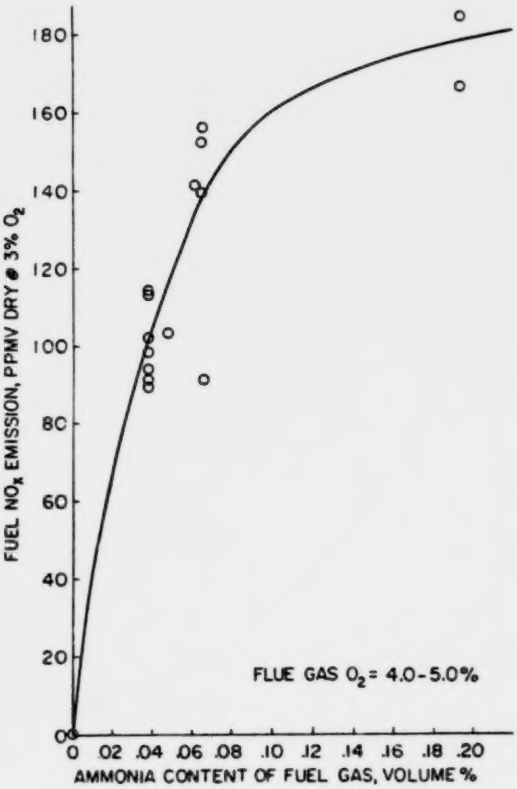

Figure 7.1-3. Fuel NO_x emissions for coal derived syngas combustion.

Source: Ref. 7-1

Used with permission of Riley Stoker Corporation from "NO_x Formation in Low- and Intermediate-Btu Coal Gas Turbulent-Diffusion Flames." February 1976.

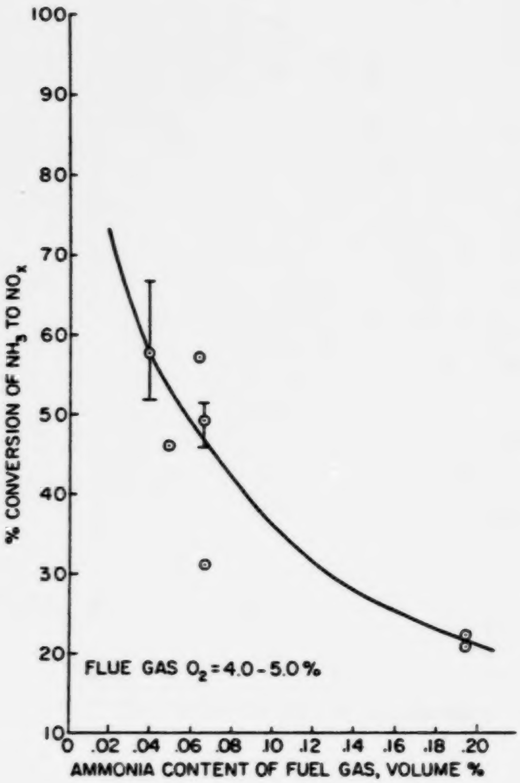

Figure 7.1-4. Conversion of NH_3 to NO_x in coal derived syngas combustion.

Source: Ref. 7-1

TABLE 7.1-2. ANALYTIC INSTRUMENTATION EQUIPMENT USED FOR IGT COMBUSTION TESTS

- Beckman 742 Polarographic Oxygen (O_2)
- Beckman Paramagnetic Oxygen (O_2)
- Beckman NDIR Methane (CH_4)
- Beckman NDIR Carbon Monoxide (CO)
- Beckman NDIR Carbon Dioxide (CO_2)
- Varian 1200 Flame Ionization Chromatograph (Total CH and C_2 to C_9)
- Beckman NDIR Nitric Oxide (NO)
- Beckman UV Nitrogen Dioxide (NO_2)
- Thermo Electron Pulsed Flourescent Sulfur Dioxide (SO_2)
- Hewlett-Packard Thermoconductivity Chromatography, Hydrogen (H), Nitrogen (N_2), Argone (A_2), CO, CO_2, C_1 to C_5, Oxygen (O_2)
- Beckman Chemiluminescent NO-NO_2

Source: Ref. 7-2

The main purpose of the IGT tests was to examine the combustion of "dirty" syngas and specifically the formation of NO_x. However, they did perform baseline tests on both clean syngas and natural gas. The results of their baseline tests are summarized in Table 7.1-3. As indicated in this table, NO_x emissions from combustion of medium-Btu gas are only 30-50 percent of those from natural gas combustion. For low-Btu gas combustion, NO_x emissions range from 13 to 34 percent of those for natural gas. These test results support the target NO_x emission control level which was based on the level of NO_x emissions from natural gas combustion.

A second part of IGT's test work involved doping the clean gases with NH_3. The results of these tests, in terms of percent conversion of NH_3 to NO_x are shown in Figures 7.1-5 and 7.1-6.

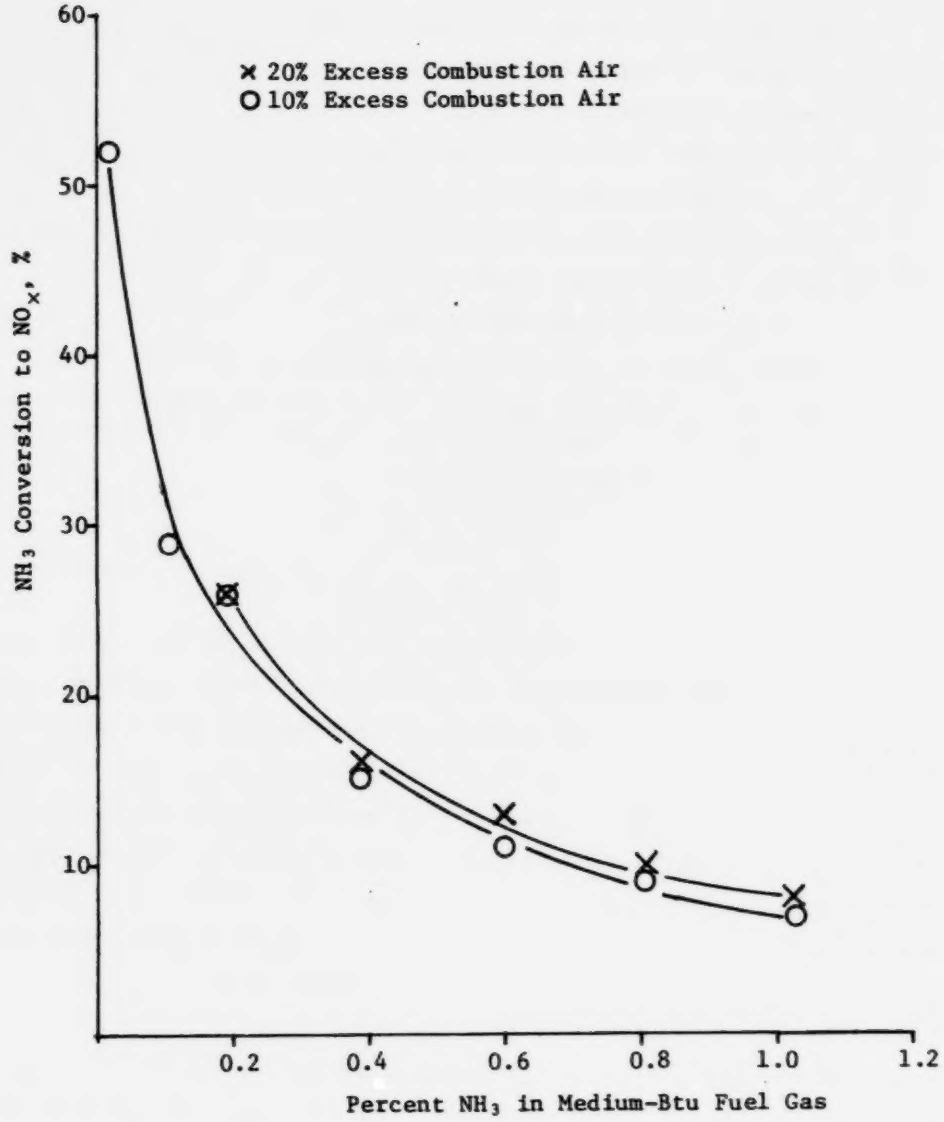

Figure 7.1-5. Conversion of NH_3 to NO_x for medium-Btu gas combustion using baffle burner.

Source: Ref. 7-2

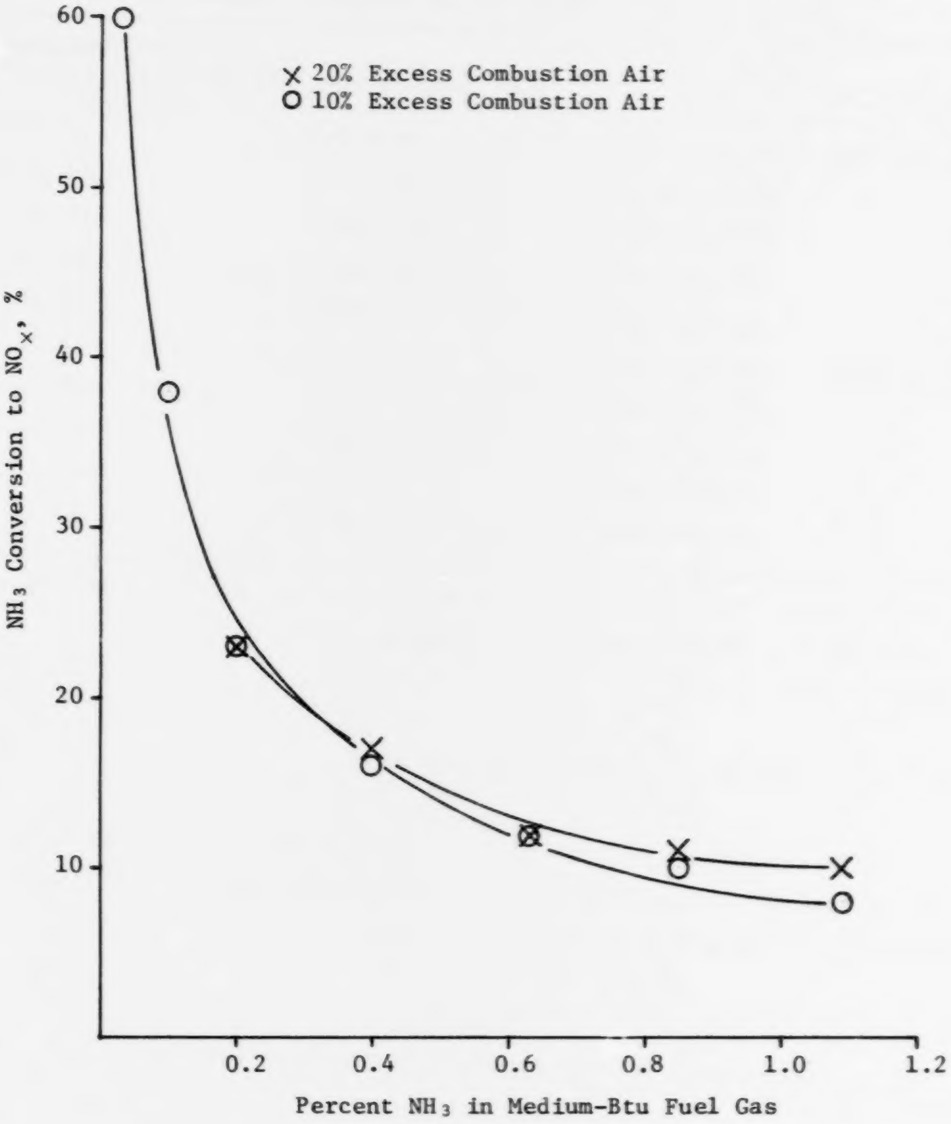

Figure 7.1-6. Conversion of NH_3 to NO_x for medium-Btu gas combustion using kiln burner.

Source: Ref. 7-2

TABLE 7.1-3. SUMMARY OF IGT BASELINE COMBUSTION TESTS*

Burner Used	Fuel	Fuel Temperature, °C(°F)	NO_x Emissions, ** ng/J (lb/10^6 Btu)
Baffle Type	Natural Gas	25 (77)	34 (0.078)
	Medium-Btu Gas	49 (120)	22 (0.051)
	Medium-Btu Gas	427 (800)	16 (0.037)
	Low-Btu Gas	49 (120)	11 (0.025)
	Low-Btu Gas	427 (800)	15 (0.035)
Kiln Type	Natural Gas	25 (77)	21 (0.049)
	Medium-Btu Gas	49 (120)	18 (0.043)
	Medium-Btu Gas	427 (800)	13 (0.030)
	Low-Btu Gas	49 (120)	17 (0.039)
	Low-Btu Gas	427 (800)	13 (0.030)

* Approximate input heat rate of 1 MW_T (3.5 x 10^6 Btu/hr) with 10% excess air.

**NO plus NO_2 (Dry, corrected to 0% excess air).

Source: Ref. 7-2

7.2 COAL LIQUIDS EMISSION SOURCE TEST DATA

This section describes eight fuel firing tests conducted to determine the combustion and emissions characteristics of coal-derived liquid fuels. The eight tests are identified as follows:

Test	Facility	Fuel
#1	Pulverized Coal-Fired B&W Utility Boiler (at Plant Mitchell)	SRC-I
#2	Cleaver-Brooks Package Boiler	EDS
#3	Scotch Laboratory Boiler	SRC-II
#4	Modified Stirling Boiler (at Alliance Research Center)	SRC-I
#5	Modified Stirling Boiler	SRC-I
#6	Solid Fuel Burning Test Facility (Combustion Engineering)	SRC-I
#7	Small Scale Laboratory Gas Turbine Combustor	SRC-II
#8	Small Scale Laboratory Gas Turbine Combustor	H-Coal

While the specific objectives of these tests varied somewhat, they all were concerned with determining the effect of different combustion parameters on air pollution emissions. Stack gas emission NO_x was monitored in all eight tests, since this is a primary concern in combusting coal-derived liquid fuels. Three of the tests (#1, #3, and #5) monitored SO_2 emissions, and five (#1, #2, #3, #4, and #5) monitored particulate concentrations in the stack gas. Table 7.2-1 summarizes the reported emissions test data. The following section provides a general description of the combustion test facilities and the emission monitoring methods used for each test.

7.2.1 Combustion Facilities and Emission Monitoring

Test #1 was conducted by Southern Company Services, Inc. to demonstrate the feasibility of burning SRC-I in a commercial utility steam boiler and to determine what modifications must be made to the boiler and auxiliary equipment to burn SRC-I fuel (Ref. 7-3, 7-4, 7-5).

This test was run on Boiler Unit No. 1 at the Georgia Power Company's Plant Mitchell, near Albany, Georgia. This is a 22.5 MW_e (rated output) Babcock & Wilcox natural circulation, pulverized coal-fired boiler. The unit is rated at 104,320 Kg of steam per hour at 6.1 MPa and 480°C (steam pressure of 900 psi and steam temperature of 900°F). It is equipped with Babcock & Wilcox E-35 pulverizers and Research Cottrell perforated-plate electrostatic precipitators.

The burners were replaced with a dual register type and minor modifications were made to the pulverizers. These modifications were required for preparation and burning of SRC-I fuel.

Emissions were measured while the unit was operating at the following output loads:

TABLE 7.2-1. EMISSIONS SOURCE TEST DATA FOR COAL DERIVED LIQUIDS

Test #	Boiler Size	Fuel Characteristics	Number or Duration of Tests	Emissions, ng/J ($lb/10^6$ Btu)	% Control Vs. Direct Coal Combustion	Control Levels Supported *	Emission Monitoring Methods	Remarks
1	22.5 MW_e electrical output	SRC-I: 35.5 MJ/kg 0.71% sulfur 0.57% ash 1.6% nitrogen	11 runs, 18 days total	NO_x: 180-200 (0.42-0.46) SO_2: 400-460 (0.93-1.07) Part: 17 (0.04)	NO_x: ~45 SO_2: ~85 Part: DNA	NO_x: Mod. Int. SO_2: Mod. Part: None	NO_x: EPA Method 7 and Chemiluminescence SO_2: EPA Method 6 and Pulsed Fluorescence Part: EPA Method 5	ESP design efficiency: 90% Average efficiency during test: 55% **
2	490 kW_T thermal output (50 hp)	EDS: 38.4 MJ/kg 0.8% sulfur 1.0% nitrogen	DNA	NO_x: 350-420 (0.81-0.98) Part: 10-12 (0.02-0.03)	NO_x: (-5)-(-55) Part: DNA	NO_x: None Part: All	NO_x: Chemiluminescence Part: EPA Method 5	
3	780 kW_T thermal output (80 hp)	SRC-II: 38.3 MJ/kg 0.17% sulfur 0.19% ash 1.08% nitrogen	47 runs	NO_x: 130-170 (0.30-0.40) SO_2: 160 (0.37) Part: 150 (0.34)	NO_x: ~(35-60) SO_2: DNA Part: DNA	NO_x: Mod. Int. SO_2: Mod. Int. Part: None	NO_x: Chemiluminescence SO_2: Flame Photometric Detector Part: EPA Method 5	
4	14 MW_T input ($47x10^6$ Btu/hr)	SRC-I: 36.3 MJ/kg 0.8% sulfur 0.29% ash 1.7% nitrogen	17 runs	NO_x: 170-240 (0.40-0.56) Part: 250-1300 (0.59-3.0)	NO_x: ~(10-50) Part: ~(65-90)	NO_x: Mod. Int. Part: None	NO_x: Chemiluminescence Part: EPA Method 5	No post combustion particulate control device
5	14 MW_T input ($47x10^6$ Btu/hr)	SRC-I: 36.3 MJ/kg 0.7% sulfur 0.4% ash 1.9% nitrogen	5 runs	NO_x: 220-270 (0.51-0.63) SO_2: 600-700 (1.4-1.6) SO_3: 4 (0.01) Part: ~900 (2)	NO_x: ~(20-35) SO_2: ~(65-70) Part: ~70	NO_x: Mod. Part: None SO_2: None	NO_x: Chemiluminescence and Nondispersive Infrared SO_2: Modified Reich Method SO_3: EPA Method 8 Part: EPA Method 5	SO_2 emissions reported are greater than the theoretical maximum possible based on fuel characteristics **
6	0.8 MW_T input ($3x10^6$ Btu/hr)	SRC-I: 34.9 MJ/kg 0.7% sulfur <0.1% ash 1.9% nitrogen	6 runs, 100 hours duration	NO_x: 260-340 (0.60-0.79)	NO_x: ~0	NO_x: Mod.	NO_x: Electrochemical Cell	
7	0.08 MW_T input ($0.3x10^6$ Btu/hr)	SRC-II: 38.9 MJ/kg 0.37% sulfur	4 runs, longest: 3 hrs	NO_x: 140 (0.32)	NO_x: ~(40-60)	NO_x: Mod. Int.	NO_x: Chemiluminescence and Electrochemical Cell	
8	0.08 MW_T input ($0.3x10^6$ Btu/hr)	H-Coal: 41.4 MJ/kg <0.002% sulfur	3 runs	NO_x: 100-130 (0.23-0.30)	NO_x: ~(50-70)	NO_x: Mod. Int.	NO_x: Chemiluminescence and Electrochemical Cell	

Source: Ref. 7-3 through 7-11.

* Mod. - Moderate control level; Int. - Intermediate control level; Str. - Stringent control level.

** Due to high excess air during combustion, NO_x values are questionable.

DNA - Data Not Available.

- Full load (22 MW_e).
- Medium load (14 MW_e), and
- Low load (7.5 MW_e).

Air pollutants were monitored by several methods. The following EPA Reference Methods were used for manual testing:

- Method 5 - For determination of particulate emissions from stationary sources,
- Method 6 - For determination of SO_2 emissions from stationary sources,
- Method 7 - For determination of nitrogen oxide emissions from stationary sources.

Nitrogen oxides were also monitored continuously and analyzed by chemiluminescence. Sulfur dioxide was also monitored by the pulsed fluorescent analysis method. In addition, grab samples were collected and analyzed by gas chromatography for the following components: CO, CO_2, SO_2, N_2, O_2, and C_1 through C_6 hydrocarbons.

Test #2 was conducted by the Exxon Research and Development Company (Ref. 7-6) to investigate the combustion properties of liquid fuel produced by the Exxon Donor Solvent Process. The test was run on a 0.49 MW_T (50 hp Cleaver-Brooks firetube package boiler which had a nominal firing rate of 57 liters/hr (15 gal/hr).

Stack particulates were measured by EPA Reference Method No. 5, while chemiluminescence was used to measure nitrogen oxide emissions.

Test #3 was conducted by KVB, Inc. (Ref. 7-7) to determine the degree to which NO_x emissions from high nitrogen coal-derived liquid fuels could be controlled by varying certain combustion parameters. The test was run on a laboratory boiler with a 0.78 MW_T (80 ph) Scotch dry-back type shell which was fired at 0.88 MW_T (3 million Btu/hr). The combustion unit had air preheat, forced air combustion, and mechanical fuel atomization.

In this test nitrogen oxides were measured by chemiluminescence, SO_2 was monitored using a photometric analyzer, and particulate emissions were determined by EPA Reference Method No. 5. Other emission test methods used were: oxygen electrolytic analysis for O_2, nondispersive infrared analysis for CO, and Bachardach Smoke Tester for smoke.

Tests #4 and #5 were conducted by the Babcock & Wilcox Company (Ref. 7-8, 7-9) to characterize certain handling, pulverization, and combustion properties of solvent refined coal (SRC-I). The tests were designed to provide information that will aid in future boiler design and retrofitting existing boilers to burn SRC-I.

These tests were run on a four-drum Stirling boiler at the B&W Alliance Research Center. This unit has a rated output of 18,000 Kg/hr (40,000 lb/hr) steam at 1.0 MPa (150 psig), which corresponds to a fuel input of about 14 MW_T (47×10^6 Btu/hr) or up to 1400 Kg/hr (3100 lb/hr) of solvent refined coal. Several modifications were made to the dual register burner, particularly the coal nozzle, of this boiler.

In Test #4, nitrogen oxide emissions were continuously monitored by chemiluminescence and particulate emissions were measured by EPA Reference Method 5. Stack opacity was measured by an optical transmissometer. Carbon monoxide was monitored using an infrared analyzer, and O_2 emissions were determined by a paramagnetic analyzer.

In Test #5, SO_2 and SO_3 emissions were also measured using EPA Reference Methods 6 and 8.

Test #6 was conducted by the CE Power Systems Division of Combustion Engineering, Inc. (Ref. 7-10) to determine if the combustion characteristics of SRC-I permit its use as a fuel in utility steam boilers. The test was run on the CE Power System's Solid Fuel Burning Test Facility, which has a rated fuel input ranging between 0.6 MW_T and 0.9 MW_T (2 and 3 x 10^6 Btu/hr). This test facility is used to simulate the combustion performance of commercial coal-fired steam boilers manufactured by CE Power Systems.

The only air pollutant monitored in Test #6 was nitrogen oxides, which was measured by the electrochemical cell method.

Tests #7 and #8 were conducted by the Westinghouse Research Laboratory (Ref. 7-11) to evaluate the use of heavy distillate coal liquids from the SRC and H-Coal processes for use in heavy duty utility turbines. The combustion test facility used was a 10 cm (4 inch) diameter scaled down combustion chamber used to simulate commercial gas turbine performance. The combustor had a fuel rate of 6.08 liters/hr (1.6 gal/hr) with a combustion air flow rate of 0.045 Kg/sec (0.1 lb/sec) at a pressure of 0.3 MPa (3 atm.). The fuel nozzle used in this test facility was of the conventional air assist atomizing type generally used with No. 2 fuels.

Test #7 was conducted to evaluate SRC-II fuels at several levels of hydroprocessing (7.4 wt percent, 10.3 wt percent and 11 wt percent hydrogen).

Test #8 was conducted to evaluate H-Coal which had been hydroprocessed at three levels (9.1 wt percent, 10.5 wt percent, and 11.7 wt percent hydrogen).

The following air pollutants were monitored in Tests #7 and #8: nitrogen oxides, carbon monoxide, carbon dioxide, unburned hydrocarbons, and smoke. Nitrogen oxides were measured both by a chemiluminescence analyzer and the electrochemical cell method. Carbon monoxide was measured by a nondispersive infrared analyzer. Unburned hydrocarbons were monitored by a flame ionization detector.

7.2.2 Comparison of Emissions

Also shown in Table 7.1-1 are comparisons of the various target emission levels to the reported emissions test data. The target control levels are summarized below:

Target Emission Control Levels for Coal-Derived Liquids	ng/J Particulates	SO_2	NO_x (as NO_2)
Moderate Control	13	520	300
Intermediate Control	13	260	220
Stringent Control	13	86	86

Particulate emissions from the EDS fuels and Test #1 for SRC-I nominally meet the stringent target emission level. However, for the other SRC-I tests (#'s 5 and 6) and the SRC-II tests, the particulate emissions are much greater than the target control levels. This is because no particulate control device was employed during these tests.

The SRC-II test data were below both the moderate and intermediate control levels for SO_2 emissions, but were greater than the stringent control level. SO_2 emissions data for SRC-I combustion were below the moderate control level in one test (#1) but above it in the other (#5). However, there is some question as to the validity of the data from Test #5 since the reported SO_2 emissions are above the theoretical maximum based on the heating value and sulfur content of the SRC-I used.

The SRC-I data were below both the moderate and intermediate control level for NO_x emissions in two of the four tests (#1 and #4). In the other two tests (#5 and #6) the range of NO_x emissions was generally only below the moderate control level. EDS combustion produced NO_x emissions which exceeded even the moderate control level. SRC-II combustion generated lower NO_x emissions than SRC-I or EDS. The test data indicate that SRC-II can meet both the moderate and intermediate control levels.

7.2.3 Discussion of Emissions Monitoring Methods

This section discusses the selection of stack monitoring methods for the three principle pollutants monitored during the combustion tests: nitrogen oxides, sulfur dioxide, and particulates. It also addresses the relative merits of the EPA Reference Method for NO_x and the other nitrogen oxide emission monitoring methods used. In the five tests in which particulate emissions were measured, the EPA Reference Method No. 5, Determination of Particulate Emissions from Stationary Sources, was used.

Sulfur dioxide emissions were monitored in three tests (#1, #3, and #5). The EPA Reference Method No. 6, a titration technique which uses barium thorin as the indicator, was used in Test #1. Test #3 used a flame photometric detector analyzer for SO_2 measurement. This technique measures SO_2 by introducing a gas sample into a hydrogen-rich flame and measuring the 3940 Å S_2 band with a narrow band optical filter. Its primary drawback is susceptibility to interference from other sulfur compounds. The photometric method was selected over the EPA Reference Method because the photometric method gives a continuous read-out. The EPA Reference Method, which necessitates accumulating the sample for several hours, is not suitable for a combustion test where a relatively small quantity of fuel is available. Test #5 used a Whittaker SS330 Analyzer which employs a modified Reich method for SO_2 determination. This method was chosen over the EPA Reference Method for the same reason as in Test #3.

Nitrogen oxide emissions were measured in all eight tests. However, only Test #1 used the EPA Reference Method No. 7. Six other tests used the chemiluminescence method, and three (Tests #6, #7, and #8) used the Electrochemical Cell method for NO_x detections.

The chemiluminescence method measures the quantity of radiant energy emitted when NO reacts with ozone to produce NO_2. This radiant energy, measured by a photomultiplier tube, is proportional to the original quantity of NO in the sample gas. Chemiluminescence analyzers can also measure NO_2 by first converting it to NO in a heated stainless steel tube and comparing the measured NO before and after conversion.

The EPA Reference Method #7 for determining NO_x is a colorimetric technique, the phenoldisulfonic acid procedure (PSD). This procedure requires that the stack gas be absorbed for several hours to collect a sufficient quantity for measurement. The chemiluminescence method was selected because it gives a continuous read-out.

Several studies have compared the chemiluminescence and colorimetric methods of NO_x determination (Ref. 7-12, 7-13, 7-14). Monteriolo and Bertolaccini (Ref. 7-12) found that the colorimetric, chemiluminescence, and electrochemical methods are equally valid in response linearity, sensitivity, and concentration limit. Bergquist (Ref. 7-13) reported that the chemiluminescence method is generally most suitable. Bourbon et al. (Ref. 7-14) found that chemiluminescence yielded theoretical results of NO_2 and NO concentrations below 1 ppm, which were more accurate than results produced by colorimetric methods.

References

7-1. Rawdon, A. H., R. A. Lisauskas and S. A. Johnson, NO_x Formation in Low and Intermediate Btu Coal Gas Turbulent-Diffusion Flames. Presented at the NO_x Control Technology Seminar, sponsored by Electric Power Research Inst. San Francisco, CA, 5-6 February 1976.

7-2. Waibel, R. T., E. S. Fleming, and D. H. Larson, Pollutant Emissions from "Dirty" Low- and Medium-Btu Gases. EPA-600/7-78-191. October 1978.

7-3. The Southern Company Services, Solvent Refined Coal Burn Test. April 1978.

7-4. Nichols, Grady B. and William J. Barrett, Evaluation of Electrostatic Precipitator During SRC Combustion Tests, EPA-600/7-78-129. July 1978.

7-5. Budden, Kenneth G., and Subhash S. Patel. Air Emissions from Combustion of Solvent Refined Coal. Final Report. Report No. EPA-600/7-79-004, EPA Contract No. 68-02-2162. Columbia, MD. Hittman Associates, Inc. January 1979.

7-6. Quinlan, C. W. and C. W. Siegmund, Combustion Properties of Coal Liquids from the Exxon Donor Solvent Process, Exxon Research and Engineering Company. Presented at American Chemical Society Symposium. Anaheim, California, March 14, 1978.

7-7. Arand, J. K., and L. J. Myzio, Combustion and Emission Evaluation of SRC Fuel Oil, A Synthetic Liquid Fuel From Coal, KVB 19900-733. Prepared for Gulf Mineral Resources by KVB, Inc. Tustin, CA, December 1977.

7-8. Babcock and Wilcox Company, Characteristics of Solvent Refined Coal: Dual Register Burner Tests. Prepared for the Electric Power Research Institute, FP-628, Research Project 1235-5, January 1978.

7-9. Babcock and Wilcox Company, Investigating Storage, Handling, and Combustion Characteristics of Solvent Refined Coal Prepared for Electric Power Research Institute, Research Project 1235-4, July 1976.

7-10. Combustion Engineering, Inc., Solvent Refined Coal Evaluation: Pulverization, Storage, and Combustion. Prepared for Electric Power Research Institute, Research Project 1235-2b, June 1976.

7-11. Cabal, A. V., M. J. Dabkowski, R. H. Heck and T. R. Stein, Utilization of Coal-Derived Liquid Fuels in a Combustion Turbine Engine. Mobile Research and Development, in American Chemical Society, preprints of papers presented at Anaheim, California, March 12-17, 1978.

7-12. Monteriolo, S. C. and M. A. Bertolaccini, Measurement of Nitrogen Dioxide in the Atmosphere. Ann. Inst. Auper. Sanita, Rome. 1973 (APTIC No. 78660).

7-13. Bergquist, K., Methods of Analysis of Nitric Oxide and Nitrogen Dioxide in the Atmosphere and in Exhaust Gases: A Literature Search. Research Institute of National Defense Report (Sweden) October 1972 (APTIC No. 79029).

7-14. Bourbon, P., J. Alary and J. C. Lepert, Determination of NO_2 and NO at Various Concentrations Using Colorimetry and Chemiluminescence. Analysis, April 1975 (APTIC No. 78217).

APPENDIX A

CAPITAL INVESTMENT REQUIREMENTS AND ANNUALIZED COSTS FOR LOW-BTU COAL GASIFICATION SYSTEMS AND LOW-BTU GAS-FIRED INDUSTRIAL BOILERS

TABLE A-1. CAPITAL INVESTMENT REQUIREMENTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
INSTALLED EQUIPMENT COSTS					
Coal Receiving and Storage	370	390	440	600	790
Gasification System	500	1,000	2,000	2,500	5,000
Gas Purification System					
Blower/Compressor	60	70	100	120	240
Quench/Cooling Towers	50	80	120	140	280
Pumps (for cooling liquor)	30	40	70	90	170
Electrostatic Precipitator	120	220	290	370	740
Separator/Evaporator	120	190	240	250	500
H_2S Removal Unit	1,330	1,380	1,800	1,970	2,300
Sulfur Recovery Unit	-	-	-	-	-
Building and Ductwork Not Costed in Above Equipment Costs (6% of EC)	160	220	320	390	640
Total Installed Equipment (EC)	2,740	3,590	5,380	6,430	10,660
INDIRECT INSTALLATION COSTS					
Engineering (10% of EC)	270	360	540	640	1,070
Construction and Field Expense (10% of EC)	270	360	540	640	1,070
Construction Fees (10% of EC)	270	360	540	640	1,070
Startup (2% of EC)	50	70	110	130	210
Total Indirect Costs (IC)	860	1,150	1,730	2,050	3,420
CONTINGENCIES (30% of EC + IC)	1,080	1,420	2,130	2,540	4,220
TOTAL TURNKEY COSTS (TTC)	4,680	6,160	9,240	11,020	18,300
WORKING CAPITAL (25% of Total Direct Operating Costs)	170	270	440	560	1,020
TOTAL CAPITAL INVESTMENT	4,850	6,430	9,680	11,580	19,320

10^3 mid-1978 dollars.

TABLE A-2. ANNUALIZED COSTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
DIRECT OPERATING COSTS					
Operating Labor (@ $12.02/hr)	210	260	320	370	580
Supervision (@ $15.63/hr)	70	70	70	70	140
Maintenance (6% of Total Turnkey Costs x 0.8)	220	300	440	530	880
Replacement Parts (included in maintenance)	-	-	-	-	-
Electricity (@ 25.8 mills/kW-hr)	20	40	80	110	210
Steam (@ $5.00/GJ)	10	30	60	80	160
Cooling Water (@ $0.05/$m^3$)	10	30	60	80	170
Process Water (@ $0.04/$m^3$)	-	-	-	-	-
Coal (@ $8.88/ton)	90	220	440	600	1190
Chemicals	20	50	110	150	300
Sulfur, Ash Disposal (@ $44.00/ton)	30	90	170	230	460
Total Direct-Operating Costs	680	1090	1750	2220	4090
OVERHEAD COSTS					
Payroll (30% of operating labor)	60	80	100	110	170
Plant (26% of labor, parts and maintenance)	130	160	220	250	420
Total Overhead Costs	190	240	320	360	590
CAPITAL CHARGES					
G&A, Local Taxes and Insurance (4% of Total Turnkey Costs)	190	250	370	440	730
Capital Recovery (11.75% of Total Turnkey Costs)	550	720	1090	1290	2150
Interest on Working Capital (@ 10%)	20	30	40	60	100
Total Capital Charges	760	1000	1500	1790	2980
TOTAL ANNUALIZED COSTS	1630	2330	3570	4370	7660
Average Gas Cost, $/GJ	9.80	5.60	4.30	3.95	3.45

10^3 Mid-1978 dollars

TABLE A-3. ANNUALIZED COSTS FOR LOW-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
CAPITAL INVESTMENT, 10^3 $					
Total Turnkey Costs (TTC)	640	1110	1740	2070	4150
Land	<10	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs, excluding fuel costs)	70	70	120	120	200
Total Capital Requirement	710	1180	1860	2190	4350
OPERATING COSTS, 10^3 $/yr					
Direct Costs					
Operating Labor (@ $12.02/hr)	110	110	210	210	320
Supervision (@ $15.63/hr)	70	70	70	70	70
Maintenance Labor (@ $14.63/hr)	30	30	60	60	130
Replacement Parts	30	40	60	70	140
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50	100
Process Water (@ $0.04/m^3)	<10	<10	<10	<10	10
Fuel	1630	2330	3570	4370	7660
Chemicals	<10	<10	10	10	10
Total Direct Operating Costs	1900	2620	4030	4840	8440
Overhead					
Payroll (30% of operating labor)	30	30	60	60	100
Plant (26% of labor and parts)	60	70	100	110	170
Total Overhead Costs	90	100	160	170	270
Capital Charges					
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80	170
Capital Recovery Factor (10.61% of TTC)	70	120	180	220	440
Interest on Working Capital (@ 10%)	10	10	10	10	20
Total Capital Charges	110	170	260	310	630
TOTAL ANNUALIZED COSTS, 10^3 $/yr	2100	2890	4450	5320	9340

Mid-1978 dollars

TABLE A-4. CAPITAL INVESTMENT REQUIREMENTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Moderate

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
INSTALLED EQUIPMENT COSTS					
Coal Receiving and Storage	370	390	440	600	790
Gasification System	500	1,000	2,000	2,500	5,000
Gas Purification System					
Blower/Compressor	60	70	100	120	240
Quench/Cooling Towers	50	80	120	140	280
Pumps (for cooling liquor)	30	40	70	90	170
Electrostatic Precipitator	100	180	240	310	620
Separator/Evaporator	90	150	170	200	400
H_2S Removal Unit	1,540	1,710	2,330	2,710	3,940
Sulfur Recovery	---	---	---	---	---
Building and Ductwork Not Costed in Above Equipment Costs (6% of EC)	170	230	350	430	730
Total Installed Equipment (EC)	2,910	3,850	5,820	7,100	12,170
INDIRECT INSTALLATION COSTS					
Engineering (10% of EC)	290	390	580	710	1,220
Construction and Field Expense (10% of EC)	290	390	580	710	1,220
Construction Fees (10% of EC)	290	390	580	710	1,220
Startup (2% of EC)	60	80	120	140	240
Total Indirect Costs (IC)	930	1,250	1,860	2,270	3,900
CONTINGENCIES (30% of EC + IC)	1,150	1,530	2,300	2,810	4,820
TOTAL TURNKEY COSTS (TTC)	4,990	6,630	9,980	12,180	20,890
WORKING CAPITAL (25% of Total Direct Operating Costs)	210	370	630	800	1,520
TOTAL CAPITAL INVESTMENT	5,200	7,000	10,610	12,980	22,410

10^3 mid-1978 dollars

TABLE A-5. ANNUALIZED COSTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Moderate
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
DIRECT OPERATING COSTS					
Operating Labor (@ $12.02/hr)	210	260	320	370	580
Supervision (@ $15.63/hr)	70	70	70	70	140
Maintenance (6% of Total Turnkey Costs x 0.8)	240	320	480	580	1000
Replacement Parts (included in maintenance)	-	-	-	-	-
Electricity (@ 25.8 mills/kW-hr)	20	60	110	150	300
Steam (@ $5.00/GJ)	<10	10	20	20	40
Cooling Water (@ $0.05/m³)	20	40	90	120	230
Process Water (@ $0.04/m³)	-	-	-	-	-
Coal (@ $18.72/ton)	160	410	810	1080	2150
Chemicals	50	120	230	310	630
Sulfur, Ash Disposal (@ $44.00/ton)	70	190	370	500	1000
Total Direct Operating Costs	840	1480	2500	3200	6070
OVERHEAD COSTS					
Payroll (30% of operating labor)	60	80	100	110	170
Plant (26% of labor, parts and maintenance)	140	170	230	270	450
Total Overhead Costs	200	250	330	380	620
CAPITAL CHARGES					
G&A, Local Taxes, and Insurance (4% of Total Turnkey Costs)	200	270	400	490	840
Capital Recovery (11.75% of Total Turnkey Costs)	590	780	1170	1430	2450
Interest on Working Capital (@ 10%)	20	40	60	80	150
Total Capital Charges	810	1090	1630	2000	3440
TOTAL ANNUALIZED COSTS	1850	2820	4460	5580	10,130
Average Gas Cost, $/GJ	11.10	6.75	5.35	5.05	4.55

10^3 Mid-1978 dollars

TABLE A-6. ANNUALIZED COSTS FOR LOW-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Moderate
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)				
	8.8(30)	22(75)	44(150)	58.6(200)	117.2(400)
CAPITAL INVESTMENT, 10^3 $					
Total Turnkey Costs (TTC)	640	1110	1740	2070	4150
Land	<10	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs, excluding fuel costs)	70	70	120	120	200
Total Capital Requirement	710	1180	1860	2190	4350
OPERATING COSTS, 10^3 $/yr					
Direct Costs					
Operating Labor (@ $12.02/hr)	110	110	210	210	320
Supervision (@ $15.63/hr)	70	70	70	70	70
Maintenance Labor (@ $14.63/hr)	30	30	60	60	130
Replacement Parts	30	40	60	70	140
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50	100
Process Water (@ $0.04/$m^3$)	<10	<10	<10	<10	10
Fuel	1850	2820	4460	5580	10,130
Chemicals	<10	<10	10	10	10
Total Direct Operating Costs	2120	3110	4920	6050	10,910
Overhead					
Payroll (30% of operating labor)	30	30	60	60	100
Plant (26% of labor and parts)	60	70	100	110	170
Total Overhead Costs	90	100	160	170	270
Capital Charges					
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80	170
Capital Recovery Factor (10.61% of TTC)	70	120	180	220	440
Interest on Working Capital (@ 10%)	10	10	10	10	20
Total Capital Charges	110	170	260	310	630
TOTAL ANNUALIZED COSTS, 10^3 $/yr	2320	3380	5340	6530	11,810

Mid-1978 dollars

TABLE A-7. CAPITAL INVESTMENT REQUIREMENTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - MEA
SO_2 Control Level - Moderate, Intermediate or Stringent

	System Capacity, MW_T (10^6 Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
INSTALLED EQUIPMENT COSTS				
Coal Receiving and Storage	370	390	440	600
Gasification System	500	1,000	2,000	2,500
Gas Purification System				
Blower/Compressor	240	290	380	530
Quench/Cooling Towers	50	80	120	140
Pumps (for cooling liquor)	30	40	70	90
Electrostatic Precipitator	100	180	240	310
Separator/Evaporator	120	190	250	280
H_2S Removal Unit	140	300	490	560
Sulfur Recovery Unit	1,640	2,170	2,710	3,000
Building and Ductwork Not Costed in Above Equipment Costs (6% of EC)	200	300	430	510
Total Installed Equipment (EC)	3,390	4,940	7,130	8,520
INDIRECT INSTALLATION COSTS				
Engineering (10% of EC)	340	490	710	850
Construction and Field Expense (10% of EC)	340	490	710	850
Construction Fees (10% of EC)	340	490	710	850
Startup (2% of EC)	70	100	140	170
Total Indirect Costs (IC)	1,090	1,570	2,270	2,720
CONTINGENCIES (30% of EC + IC)	1,340	1,950	2,820	3,370
TOTAL TURNKEY COSTS (TTC)	5,820	8,460	12,220	14,610
WORKING CAPITAL (25% of Total Direct Operating Costs)	230	420	700	990
TOTAL CAPITAL INVESTMENT	6,050	6,880	12,920	15,600

10^3 mid-1978 dollars

TABLE A-8. ANNUALIZED COSTS FOR LOW-BTU WELLMAN-GALUSHA GASIFICATION SYSTEMS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - MEA
SO_2 Control Level - Moderate, Intermediate or Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
DIRECT OPERATING COSTS				
Operating Labor (@ \$12.02/hr)	210	260	320	370
Supervision (@ \$15.63/hr)	70	70	70	70
Maintenance (6% of Total Turnkey Costs x 0.8)	280	410	590	700
Replacement Parts (included in maintenance)	---	---	---	---
Electricity (@ 25.8 mills/kW-hr)	70	160	330	440
Steam (@ \$5.00/GJ)	50	120	230	310
Cooling Water (@ \$0.05/m^3)	20	50	90	120
Process Water (@ \$0.04/m^3)	---	---	---	---
Coal (@ \$18.72/ton)	160	410	810	1080
Chemicals	10	30	60	80
Sulfur, Ash Disposal (@ \$44.00/ton)	60	160	320	420
Total Direct Operating Costs	930	1670	2820	3590
OVERHEAD COSTS				
Payroll (30% of operating labor)	60	80	100	110
Plant (26% of labor, parts and maintenance)	150	190	250	300
Total Overhead Costs	210	270	350	410
CAPITAL CHARGES				
G&A, Local Taxes, and Insurance (4% of Total Turnkey Costs)	230	340	490	580
Capital Recovery (11.75% of Total Turnkey Costs)	680	990	1440	1720
Interest on Working Capital (@ 10%)	20	40	70	100
Total Capital Charges	930	1370	2000	2400
TOTAL ANNUALIZED COSTS	2070	3310	5170	6400
Average Gas Cost, \$/GJ	12.45	7.95	6.20	5.75

10^3 mid-1978 dollars

TABLE A-9. ANNUALIZED COSTS FOR LOW-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - MEA
SO_2 Control Level - Moderate, Intermediate or Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 \$				
Total Turnkey Costs (TTC)	640	1,110	1,740	2,070
Land	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs, excluding fuel costs)	70	70	120	120
Total Capital Requirement	750	1,180	1,860	2,190
OPERATING COSTS, 10^3 \$/yr				
Direct Costs				
Operating Labor (@ \$12.02/hr)	110	110	210	210
Supervision (@ \$15.63/hr)	70	70	70	70
Maintenance Labor (@ \$14.63/hr)	30	30	60	60
Replacement Parts	30	40	60	70
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50
Process Water (@ \$0.04/m^3)	<10	<10	<10	<10
Fuel	2,070	3,310	5,170	6,400
Chemicals	<10	<10	10	10
Total Direct Operating Costs	2,340	3,600	5,630	6,870
Overhead				
Payroll (30% of operating labor)	30	30	60	60
Plant (26% of labor and parts)	60	70	100	110
Total Overhead Costs	90	100	160	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80
Capital Recovery Factor (10.61% of TTC)	70	120	180	220
Interest on Working Capital (@ 10%)	10	10	10	10
Total Capital Charges	110	170	260	310
TOTAL ANNUALIZED COSTS, 10^3 \$/yr	2,540	3,870	6,050	7,350

Mid-1978 dollars

APPENDIX B

CAPITAL INVESTMENT REQUIREMENTS AND ANNUALIZED COSTS FOR MEDIUM-BTU COAL GASIFICATION SYSTEMS AND MEDIUM-BTU GAS-FIRED INDUSTRIAL BOILERS

TABLE B-1. ESTIMATED INSTALLED EQUIPMENT COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS[a]

Process Area	Low Sulfur Western Coal		High Sulfur Eastern Coal	
	Stretford	Rectisol	Stretford	Rectisol
Coal Storage and Reclaiming	17	17	16	16
Coal Preparation	20	20	19	19
Coal Feed Gasification Raw Gas Quench	162	162	162	162
Acid Gas Removal	11	136	33	169
Product Gas Drying	0.9	0.9	0.9	0.9
Sour Water Stripping, Ammonia Recovery, and Bio-Oxidation	22	22	22	22
Sulfur Recovery	-	16	-	40
Solids Disposal	3.5	3.5	5.6	5.5
Steam and Utility Systems	228	233	324	316
Plant Water Systems	26	26	26	26
Oxygen Plant	77	77	95	95
General Facilities	96	96	96	96
TOTAL INSTALLED EQUIPMENT COSTS	663.4	809.4	799.5	967.4

[a] 3060 MW_T (250×10^9 Btu/day) capacity (gas output); mid-1978 dollars; 10^6 dollars.

TABLE B-2. ESTIMATED TOTAL CAPITAL REQUIREMENTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEMS[a]

	Low Sulfur western coal		High sulfur eastern coal	
	Stretford	Rectisol	Stretford	Rectisol
Total Engineered Equipment Costs (EC) (including direct installation costs)	663.4	809.4	799.5	967.4
Installation Costs, indirect				
Engineering and Fee (10% of EC)	66.3	80.9	80.0	96.7
Construction and Field Expense (included in Equipment Costs)				
Contingencies (30% of EC)	218.9	267.1	263.9	319.2
Allowance for Funds Used During Construction (EC x Average Spending Period (1.75 yrs) x 10%)	116.1	141.6	139.9	169.3
Paid-up Royalties	1.1	1.1	1.1	1.1
Startup Costs (20% of total gross annual operating costs)	23.1	24.4	38.4	40.8
Land	0.9	0.9	0.9	0.9
Working Capital (25% total direct operating costs)	26.0	27.7	45.0	47.3
Total Capital Requirement	1115.8	1353.1	1368.7	1642.7

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[a] 3060 MW_T (250 x 10^9 Btu/day) capacity (gas output); mid-1978 dollars; 10^6 dollars.

TABLE B-3. ESTIMATED MAINTENANCE COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock — Low sulfur western
Acid Gas Removal Unit — Stretford

Unit	Unit Cost	Maintenance Factor, %	Maintenance Cost
Coal Handling and Reclaiming	24	6	1.4
Coal Preparation	29	6	1.7
Coal Feed Gasification Raw Gas Quench	232	5	11.6
Acid Gas Removal	16	3	0.5
Product Gas Drying	1	3	0.03
Liquid Effluent Treatment	31	3	0.9
Sulfur Recovery	-	3	-
Solids Disposal	5	3	0.2
Steam and Utilities Systems	326	1	3.3
Plant Water Systems	37	3	1.1
Oxygen Plant	110	3	3.3
General Facilities	137	1	1.4
TOTAL	948		25.4
Maintenance Labor (@ 60% of Total Maintenance)			15.2
Maintenance Supplies (@ 40% of Total Maintenance)			10.8

[a] 3060 MW_T (250×10^9 Btu/day) gas output capacity; 10^6 mid-1978 dollars.

TABLE B-4. ANNUALIZED COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
Operating Factor - 90%

Direct Costs			
Operating Labor (@ \$12.02/hr)	6.3		
Maintenance Labor (60% of Total Maintenance)	15.2		
Supervision (20% of Operating and Maintenance Labor)	4.3		
Maintenance Materials (40% of Total Maintenance)	10.8		
Replacement Parts (Included in Maintenance Materials)	-		
Operating Supplies (30% of Operating Labor)	1.9		
Purchased Water (@ \$0.032/m^3)	0.1		
Coal (@ \$8.88/ton)	58.3		
Chemicals and Catalysts	2.3		
Solids Disposal (@ \$11/ton)	4.7		
Total Direct Costs		103.9	
Overhead Costs			
Payroll (30% of Operating Labor)	1.9		
Plant (26% of Labor, Materials, and Maintenance)	9.5		
Total Overhead Costs		11.4	
By-Product Credits			
Sulfur (@ \$29/ton)	0.8		
Ammonia (@ \$110/ton)	5.2		
Naphtha (@ \$93/m^3)	-		
Phenols (@ \$46/m^3)	1.1		
Tars (@ \$46/m^3)	8.7		
Oils (@ \$79/m^3)	8.7		
Total By-Product Credits		(24.5)	
Capital Related Charges			
G&A, Local Taxes, and Insurance (4% of Total Capital Requirement Excluding Start-up Costs, Land, and Working Capital)	42.6		
Capital Recovery (11.75% of Total Capital Requirement Excluding Land and Working Capital)	127.9		
Interest on Working Capital (@ 10%)	2.6		
Total Capital Related Charges		173.1	
TOTAL ANNUALIZED COSTS			263.9
Average Gas Costs, \$/GJ (\$/10^6 Btu)			3.05 (3.20)
Average Gas Costs[b], \$/GJ \$/10^6 Btu)			4.30 (4.55)

[a] 3060 MW_T (250 x 10^9 Btu/day) output gas capacity; 10^6 mid-1978 dollars.

[b] At 60% operating factor.

TABLE B-5. ESTIMATED MAINTENANCE COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock — Low sulfur western
Acid Gas Removal Unit — Rectisol

Unit	Unit Cost	Maintenance Factor, %	Maintenance Cost
Coal Handling and Reclaiming	24	6	1.4
Coal Preparation	29	6	1.7
Coal Feed Gasification Raw Gas Quench	232	5	11.6
Acid Gas Removal	194	3	5.8
Product Gas Drying	1	3	0.03
Liquid Effluent Treatment	31	3	0.9
Sulfur Recovery	23	3	0.7
Solids Disposal	5	3	0.2
Steam and Utilities Systems	333	1	3.3
Plant Water Systems	37	3	1.1
Oxygen Plant	110	3	3.3
General Facilities	137	1	1.4
TOTAL	1156		31.4
Maintenance Labor (@ 60% of Total Maintenance)			18.8
Maintenance Supplies (@ 40% of Total Maintenance)			12.6

[a] 3060 MW_T (250 x 10^9 Btu/day) gas output capacity; 10^6 mid-1978 dollars.

TABLE B-6. ANNUALIZED COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Rectisol
Operating Factor - 90%

Direct Costs			
Operating Labor (@ \$12.02/hr)	6.3		
Maintenance Labor (60% of Total Maintenance)	18.8		
Supervision (20% of Operating and Maintenance Labor)	5.0		
Maintenance Materials (40% of Total Maintenance)	12.6		
Replacement Parts (Included in Maintenance Materials)	-		
Operating Supplies (30% of Operating Labor)	1.9		
Purchased Water (@ \$0.032/m^3)	0.1		
Coal (@ \$8.88/ton)	58.9		
Chemicals and Catalysts	2.3		
Solids Disposal (@ \$11/ton)	4.8		
Total Direct Costs		110.7	
Overhead Costs			
Payroll (30% of Operating Labor)	1.9		
Plant (26% of Labor, Materials, and Maintenance)	9.4		
Total Overhead Costs		11.3	
By-Product Credits			
Sulfur (@ \$29/ton)	0.8		
Ammonia (@ \$110/ton)	5.2		
Naphtha (@ \$93/m^3)	6.6		
Phenols (@ \$46/m^3)	1.1		
Tars (@ \$46/m^3)	8.7		
Oils (@ \$79/m^3)	8.7		
Total By-Product Credits		(31.1)	
Capital Related Charges			
G&A, Local Taxes and Insurance (4% of Total Capital Requirement Excluding Start-up Costs, Land, and Working Capital)	52.0		
Capital Recovery (11.75% of Total Capital Requirement Excluding Land and Working Capital)	155.6		
Interest on Working Capital (@ 10%)	2.8		
Total Capital Related Charges		210.4	
TOTAL ANNUALIZED COSTS			301.3
Average Gas Costs, \$/GJ (\$/10^6 Btu)			3.45 (3.65)
Average Gas Costs[b], \$/GJ (\$/10^6 Btu)			5.00 (5.30)

[a] 3060 MW_T (250 x 10^9 Btu/day) output gas capacity; 10^6 mid-1978 dollars.

[b] At 60% operating factor.

TABLE B-7. ESTIMATED MAINTENANCE COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock	High sulfur eastern
Acid Gas Removal Unit	Stretford

Unit	Unit Cost	Maintenance Factor, %	Maintenance Cost
Coal Handling and Reclaiming	23	6	1.4
Coal Preparation	27	6	1.6
Coal Feed	232	5	11.6
Gasification			
Raw Gas Quench			
Acid Gas Removal	47	3	1.4
Product Gas Drying	1	3	0.03
Liquid Effluent Treatment	31	3	0.9
Sulfur Recovery	-	3	-
Solids Disposal	8	3	0.2
Steam and Utilities Systems	463	1	4.6
Plant Water Systems	37	3	1.1
Oxygen Plant	136	3	4.1
General Facilities	137	1	1.4
TOTAL	1142		28.3
Maintenance Labor (@ 60% of Total Maintenance)			17.0
Maintenance Supplies (@ 40% of Total Maintenance)			11.3

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[a] 3060 MW_T (250 x 10^9 Btu/day) gas ouput capacity; 10^6 mid-1978 dollars.

TABLE B-8. ANNUALIZED COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Stretford
Operating Factor - 90%

Direct Costs			
Operating Labor (@ $12.02/hr)	6.3		
Maintenance Labor (60% of Total Maintenance)	17.0		
Supervision (20% of Operating and Maintenance Labor)	4.7		
Maintenance Materials (40% of Total Maintenance)	11.3		
Replacement Parts (Included in Maintenance Materials)	-		
Operating Supplies (30% of Operating Labor)	1.9		
Purchased Water (@ $0.032/m^3)	0.1		
Coal (@ $18.72/ton)	116.6		
Chemicals and Catalysts	6.4		
Solids Disposal (@ $11/ton)	15.6		
Total Direct Costs		179.9	
Overhead Costs			
Payroll (30% of Operating Labor)	1.9		
Plant (26% of Labor, Materials, and Maintenance)	10.2		
Total Overhead Costs		12.1	
By-Product Credits			
Sulfur (@ $29/ton)	4.3		
Ammonia (@ $110/ton)	3.3		
Naphtha (@ $93/m^3)	-		
Phenols (@ $46/m^3)	0.7		
Tars (@ $46/m^3)	5.8		
Oils (@ $79/m^3)	1.2		
Total By-Product Credits		(15.3)	
Capital Related Charges			
G&A, Local Taxes and Insurance (4% of Total Capital Requirement Excluding Start-up Costs, Land, and Working Capital)	51.4		
Capital Recovery (11.75% of Total Capital Requirement Excluding Land and Working Capital)	155.4		
Interest on Working Capital (@ 10%)	4.5		
Total Capital Related Charges		211.3	
TOTAL ANNUALIZED COSTS			388.0
Average Gas Costs, $/GJ ($/10^6 Btu)			4.45 (4.70)
Average Gas Costs[b], $/GJ ($/10^6 Btu)			6.00 (6.35)

[a] 3060 MW$_T$ (250 x 10^9 Btu/day) output gas capacity; 10^6 mid-1978 dollars.

[b] At 60% operating factor.

TABLE B-9. ESTIMATED MAINTENANCE COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock — High sulfur eastern

Acid Gas Removal Unit — Rectisol

Unit	Unit Cost	Maintenance Factor, %	Maintenance Cost
Coal Handling and Reclaiming	23	6	1.4
Coal Preparation	27	6	1.6
Coal Feed Gasification Raw Gas Quench	232	5	11.6
Acid Gas Removal	242	3	7.3
Product Gas Drying	1	3	0.03
Liquid Effluent Treatment	31	3	0.9
Sulfur Recovery	57	3	1.7
Solids Disposal	8	3	0.2
Steam and Utilities Systems	452	1	4.5
Plant Water Systems	37	3	1.1
Oxygen Plant	136	3	4.1
General Facilities	137	1	1.4
TOTAL	1383		35.8
Maintenance Labor (@ 60% of Total Maintenance)			21.5
Maintenance Supplies (@ 40% of Total Maintenance)			14.3

[a] 3060 MW_T (250 x 10^9 Btu/day) gas output capacity; 10^6 mid-1978 dollars.

TABLE B-10. ANNUALIZED COSTS FOR LURGI MEDIUM-BTU GASIFICATION SYSTEM[a]

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Rectisol
Operating Factor - 90%

Direct Costs			
Operating Labor (@ \$12.02/hr)	6.3		
Maintenance Labor (60% of Total Maintenance)	21.5		
Supervision (20% of Operating and Maintenance Labor)	6.6		
Maintenance Materials (40% of Total Maintenance)	14.3		
Replacement Parts (Included in Maintenance Materials)	-		
Operating Supplies (30% of Operating Labor)	1.9		
Purchased Water (@ \$0.032/m^3)	0.1		
Coal (@ \$18.72/ton)	116.0		
Chemicals and Catalysts	6.4		
Solids Disposal (@ \$11/ton)	16.1		
Total Direct Costs		189.2	
Overhead Costs			
Payroll (30% of Operating Labor)	1.9		
Plant (26% of Labor, Materials, and Maintenance)	12.7		
Total Overhead Costs		14.6	
By-Product Credits			
Sulfur (@ \$29/ton)	4.3		
Ammonia (@ \$110/ton)	3.3		
Naphtha (@ \$93/m^3)	4.4		
Phenols (@ \$46/m^3)	0.7		
Tars (@ \$46/m^3)	5.8		
Oils (@ \$79/m^3)	1.2		
Total By-Product Credits		(19.7)	
Capital Related Charges			
G&A, Local Taxes and Insurance (4% of Total Capital Requirement Excluding Start-up Costs, Land, and Working Capital)	62.1		
Capital Recovery (11.75% of Total Capital Requirement Excluding Land and Working Capital)	187.4		
Interest on Working Capital (@ 10%)	4.7		
Total Capital Related Charges		254.2	
TOTAL ANNUALIZED COSTS			438.3
Average Gas Costs, \$/GJ (\$/10^6 Btu)			5.05 (5.35)
Average Gas Costs[b], \$/GJ (\$/10^6 Btu)			6.90 (7.30)

[a] 3060 MW_T (250 x 10^9 Btu/day) output gas capacity; 10^6 mid-1978 dollars.

[b] At 60% operating factor.

TABLE B-11. ANNUALIZED COSTS FOR MEDIUM-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 \$				
Total Turnkey Costs (TTC)	640	110	1,740	2,070
Land	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs)	250	520	1,010	1,310
Total Capital Requirement	890	630	2,750	3,380
OPERATING COSTS, 10^3 \$/yr				
Direct Costs				
Operating Labor (@ \$12.02/hr)	110	110	210	210
Supervision (@ \$15.63/hr)	70	70	70	70
Maintenance Labor (@ \$14.63/hr)	30	30	60	60
Replacement Parts	30	40	60	70
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50
Process Water (@ \$0.04/m^3)	<10	<10	<10	<10
Fuel (@ \$4.30/GJ)	720	1,790	3,580	4,770
Chemicals	<10	<10	10	10
Total Direct Operating Costs	990	2,080	4,040	5,240
Overhead				
Payroll (30% of operating labor)	30	30	60	60
Plant (26% of labor and parts)	60	70	100	110
Total Overhead Costs	90	100	160	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80
Capital Recovery Factor (10.61% of TTC)	70	120	180	220
Interest on Working Capital (@ 10%)	30	50	100	130
Total Capital Charges	130	210	350	430
TOTAL ANNUALIZED COSTS, 10^3 \$/yr	1,210	2,390	4,550	5,840

Mid-1978 dollars.

TABLE B-12. ANNUALIZED COSTS FOR MEDIUM-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - Low Sulfur Western
Acid Gas Removal Unit - Rectisol
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 $				
Total Turnkey Costs (TTC)	640	1,110	1,740	2,070
Land	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs)	280	590	1,160	1,500
Total Capital Requirement	920	1,700	2,900	3,570
OPERATING COSTS, 10^3 $/yr				
Direct Costs				
Operating Labor (@ $12.02/hr)	110	110	210	210
Supervision (@ $15.63/hr)	70	70	70	70
Maintenance Labor (@ $14.63/hr)	30	30	60	60
Replacement Parts	30	40	60	70
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50
Process Water (@ $0.04/$m^3$)	<10	<10	<10	<10
Fuel (@ $5.00/GJ)	830	2,080	4,160	5,540
Chemicals	<10	<10	10	10
Total Direct Operating Costs	1,100	2,370	4,620	6,010
Overhead				
Payroll (30% of operating labor)	30	30	60	60
Plant (26% of labor and parts)	60	70	100	110
Total Overhead Costs	90	100	160	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80
Capital Recovery Factor (10.61% of TTC)	70	120	180	220
Interest on Working Capital (@ 10%)	30	60	120	150
Total Capital Charges	130	220	370	450
TOTAL ANNUALIZED COSTS, 10^3 $/yr	1,320	2,690	5,150	6,630

Mid-1978 dollars.

TABLE B-13. ANNUALIZED COSTS FOR MEDIUM-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Stretford
SO_2 Control Level - Intermediate
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 $				
Total Turnkey Costs (TTC)	640	1,110	1,740	2,070
Land	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs)	320	700	1,370	1,780
Total Capital Requirement	960	1,810	3,100	3,850
OPERATING COSTS, 10^3 $/yr				
Direct Costs				
Operating Labor (@ $12.02/hr)	110	110	210	210
Supervision (@ $15.63/hr)	70	70	70	70
Maintenance Labor (@ $14.63/hr)	30	30	60	60
Replacement Parts	30	40	60	70
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50
Process Water (@ $0.04/$m^3$)	<10	<10	<10	<10
Fuel (@ $6.00/GJ)	1,000	2,500	5,000	6,650
Chemicals	<10	<10	10	10
Total Direct Operating Costs	1,270	2,790	5,460	7,120
Overhead				
Payroll (30% of operating labor)	30	30	60	60
Plant (26% of labor and parts)	60	70	100	110
Total Overhead Costs	90	100	160	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80
Capital Recovery Factor (10.61% of TTC)	70	120	180	220
Interest on Working Capital (@ 10%)	30	70	140	180
Total Capital Charges	130	230	390	480
TOTAL ANNUALIZED COSTS, 10^3 $/yr	1,490	3,120	6,010	7,770

Mid-1978 dollars.

TABLE B-14. ANNUALIZED COSTS FOR MEDIUM-BTU GAS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - High Sulfur Eastern
Acid Gas Removal Unit - Rectisol
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 $				
Total Turnkey Costs (TTC)	640	1,110	1,740	2,070
Land	<10	<10	<10	<10
Working Capital (25% of Total Direct Operating Costs)	360	790	1,550	2,030
Total Capital Requirement	1,000	1,900	3,290	4,100
OPERATING COSTS, 10^3 $/yr				
Direct Costs				
Operating Labor (@ $12.02/hr)	110	110	210	210
Supervision (@ $15.63/hr)	70	70	70	70
Maintenance Labor (@ $14.63/hr)	30	30	60	60
Replacement Parts	30	40	60	70
Electricity (@ 25.8 mills/kW-hr)	30	40	50	50
Process Water (@ $0.04/$m^3$)	<10	<10	<10	<10
Fuel (@ $6.90/GJ)	1,150	2,870	5,740	7,650
Chemicals	<10	<10	10	10
Total Direct Operating Costs	1,420	3,160	6,200	8,120
Overhead				
Payroll (30% of operating labor)	30	30	60	60
Plant (26% of labor and parts)	60	70	100	110
Total Overhead Costs	90	100	160	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	30	40	70	80
Capital Recovery Factor (10.61% of TTC)	70	120	180	220
Interest on Working Capital (@ 10%)	40	80	160	200
Total Capital Charges	140	240	410	500
TOTAL ANNUALIZED COSTS, 10^3 $/yr	1,650	3,500	6,770	8,790

Mid-1978 dollars.

APPENDIX C

CAPITAL INVESTMENT REQUIREMENTS AND ANNUALIZED COSTS FOR COAL LIQUEFACTION SYSTEMS AND COAL LIQUIDS-FIRED INDUSTRIAL BOILERS

TABLE C-1. TOTAL CAPITAL REQUIREMENT FOR SRC-I PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	7050 MW_T (577 x 10^9 Btu/day)
SO_2 Control Level	Moderate
Installed Equipment Costs	
Coal preparation	90
Hydrogenation and hydrogen recycle	221
Fractionation	27
Hydrogen plant	342
Filtration	192
Product solidification	28
Gas and secondary recovery	90
Offsites and wastewater treatment	110
General Facilities	86
Total Installed Equipment (EC)	1,186
Engineering and fee (included in direct costs)	-
Construction and field expense (included in direct costs)	-
Contingencies (30% of EC)	356
Start-up Costs (20% of total gross annual operating costs)	59
Allowance for funds used during construction (EC x 1.75 x 0.10)	208
Land	2
Working capital (25% of total direct annual operating costs)	73
Total Indirect Costs	698
TOTAL CAPITAL REQUIREMENT	1,884

10^6 mid-1978 dollars

TABLE C-2. ANNUALIZED COSTS FOR SRC-I PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	7050 MW_T (577 x 10^9 Btu/day)
SO_2 Control Level	Moderate
Operating Factor	80%
Direct Costs	
Manpower	31
Repair materials and other	48
Electricity (@ 25.8 mills/kW-h)	16
Purchased water (@ \$0.032/m^3)	<1
Catalysts and Chemicals	6
Coal (@ \$18.72/ton)	163
Total Direct Costs	264
Overhead	
Payroll (30% of labor)	9
Plant (26% of labor, parts and maintenance)	20
Total Overhead Costs	29
General and Administrative Costs, Local Taxes and Insurance (4% of total capital requirement excluding land and working capital)	72
By-Product Credits	(9)
Capital Recovery Factor (11.75% of total capital requirement excluding land and working capital)	213
Interest on working capital	7
TOTAL ANNUALIZED COSTS	576
Average Product Cost, \$/GJ	3.25

10^6 mid-1978 dollars

TABLE C-3. ANNUALIZED COSTS FOR SRC (SOLID)-FIRED INDUSTRIAL BOILER

Boiler Size	58.6 MW_T (200 x 10^6 Btu/hr)		
Coal Feedstock	High sulfur bituminous		
SO_2 Control Level	Moderate		
Operating Factor	60%		
CAPITAL INVESTMENT, $10^3\$$			
Total Turnkey Costs (TTC)	10,199		
Land	2		
Working capital (25% of total direct operating costs)	1,213		
Total Capital Requirement	11,414		
OPERATING COSTS, $10^3\$/yr$			
Direct Costs			
Direct labor (@ $12.02/hr)	421		
Supervision (@ $15.63/hr)	137		
Maintenance labor (@ $14.63/hr)	192		
Replacement parts	250		
Electricity (@ 25.8 mills/kWhr)	238		
Process water (@ $\$0.04/m^3$)	3		
Fuel (@ $3.25/GJ)	3,604		
Chemicals	8		
Total Direct Operating Costs		4,853	
Overhead			
Payroll (30% of labor)	226		
Plant (26% of labor and parts)	260		
Total Overhead Costs		486	
Capital Charges			
G&A, local taxes, and insurance (4% of TTC)	408		
Capital recovery factor (10.14% of TTC)	1,034		
Interest on working capital (@ 10%)	121		
Total Capital Charges		1,563	
TOTAL ANNUALIZED COSTS, $10^3\$/yr$			6,802

Mid-1978 dollars.

TABLE C-4. TOTAL CAPITAL REQUIREMENTS FOR EXXON DONOR SOLVENT PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	4000 MW_T (328 x 10^9 Btu/day)
SO_2 Control Level	Intermediate*
Installed Equipment Costs	
Liquefaction	218
Solvent hydrogenation	74
Flexicoker	145
Hydrogen recovery and generation	218
Gas and wastewater treatment	44
Product recovery	7
Offsites	224
Total Equipment Costs (EC)	930
Engineering and fee (10% of EC)	93
Construction and field expense (included in direct costs)	
Contingencies (30% of EC + Engineering + Fee)	307
Start-up costs (20% of total gross annual operating costs	53
Allowance for funds used during construction (EC x 1.75 x 0.10)	163
Land	2
Working capital (25% of total direct annual operating costs)	66
Total Indirect Costs	684
TOTAL CAPITAL REQUIREMENT	1,614

*Additional hydroprocessing will permit attainment of stringent control levels. Additional costs incurred for hydroprocessing cannot be determined and are assumed to be negligible.

10^6 mid-1978 dollars

TABLE C-5. ANNUALIZED COSTS FOR EXXON DONOR SOLVENT PROCESS

Coal Feedstock	High Sulfur Bituminous
Capacity	4000 MW (328 x 10^9 Btu/day)
SO_2 Control Level	Intermediate*
Operating Factor	80%

Direct Costs	
Manpower	31
Repair materials and other	48
Electricity (@ 25.8 mills/kW-h)	32
Purchased water (@ $0.032/$m^3$)	<1
Catalysts and Chemicals	6
Coal (@ $18.72/ton)	119
Total Direct Costs	236
Overhead	
Payroll (30% of labor)	9
Plant (26% of labor, parts and maintenance)	20
Total Overhead Costs	29
General and Administrative Costs, Local Taxes and Insurance (4% of total capital requirements excluding land and working capital)	62
By-Product Credits	(16)
Capital Recovery Factor (11.75% of total capital requirement excluding land and working capital)	213
Interest on working capital	7
TOTAL ANNUALIZED COSTS	576
Average Product Cost, $/GJ	4.95

*Additional hydroprocessing will permit attainment of stringent control levels. Additional costs incurred for hydroprocessing cannot be determined and are assumed to be negligible.

10^6 mid-1978 dollars

TABLE C-6. ANNUALIZED COSTS FOR EDS LIQUIDS-FIRED INDUSTRIAL BOILERS

Coal Feedstock - High Sulfur Eastern
SO_2 Control Level - Stringent
Operating Factor - 60%

	System Capacity, MW_T (10^6 Btu/hr)			
	8.8(30)	22(75)	44(150)	58.6(200)
CAPITAL INVESTMENT, 10^3 $				
Total Turnkey Costs (TTC)	636	1,102	1,729	2,055
Land	2	2	2	2
Working Capital (25% of Total Direct Operating Costs	273	587	1,145	1,490
Total Capital Requirement	911	1,691	2,876	3,547
OPERATING COSTS, 10^3 $/yr				
Direct Costs				
Operating Labor (@ $12.02/hr)	105	105	210	210
Supervision (@ $15.63/hr)	68	68	68	68
Maintenance Labor (@ $14.63/hr)	32	32	64	64
Replacement Parts	30	40	60	71
Electricity (@ 25.8 mills/kW-hr)	29	40	47	50
Process Water (@ $0.04/$m^3$)	<1	1	2	3
Fuel (@ $4.95/GJ)	824	2,060	4,121	5,489
Chemicals	2	3	6	7
Total Direct Operating Costs	1,090	2,349	4,578	5,962
Overhead				
Payroll (30% of operating labor)	32	32	63	63
Plant (26% of labor and parts)	61	64	104	107
Total Overhead Costs	93	96	167	170
Capital Charges				
G&A, Local Taxes, and Insurance (4% of TTC)	25	44	69	82
Capital Recovery Factor (10.61% of TTC)	67	117	183	218
Interest on Working Capital (@ 10%)	27	59	115	149
Total Capital Charges	119	220	367	449
TOTAL ANNUALIZED COSTS, 10^3 $/yr	1,302	2,665	5,112	6,581

Mid-1978 dollars

APPENDIX D

ENERGY REQUIREMENTS FOR LOW-BTU GASIFICATION SYSTEMS - EXAMPLE CALCULATION

APPENDIX D

ENERGY REQUIREMENTS FOR LOW-BTU GASIFICATION SYSTEMS - EXAMPLE CALCULATION

In order to illustrate the procedures used to calculate the low-Btu gasification energy impact data contained in Table 5.1-2, an example calculation is presented below. The case considered in the example is gasification of low sulfur western coal and use of the Stretford process for H_2S removal. Also included as an example is the calculation of the energy requirements for the acid gas removal (MEA process) and sulfur recovery (Claus/SCOT processes) units for the high sulfur coal cases.

D.1 LOW SULFUR COAL CASE - EXAMPLE CALCULATION

The basis for this example calculation is a plant producing 8.79 MW_T (30×10^6 Btu/hr) of low-Btu gas from low sulfur western coal. Based on the desired plant output and raw gas heating value, the gas flow rate from the gasifier is calculated to be 1.60 m^3/s. To produce this gas requires a coal feed rate of 0.528 kg/s. Since the coal has a higher heating value of 22.3 MJ/kg, the energy lost to conversion inefficiencies is approximately

$$22.3 \times 0.528 - 8.79 = 2.98 \ MW_T$$

Tar and oil formation is assumed to be 10 percent of the coal feed. Therefore, about 0.053 kg/s of tars and oils are produced. At 99.5 percent recovery of these by-products and an estimated higher heating value of 37.2 MJ/kg, the energy content of the by-products recovered is

$$0.053 \times 0.995 \times 37.2 = \underline{1.96\ \text{MW}_T}.$$

The rest of the gasification inefficiencies are attributable to conversion losses and are calculated to be

$$2.98 - 1.96 = \underline{1.02\ \text{MW}_T}.$$

Electricity is required for the gasifier inlet air blower and the pumps that circulate the gasifier cooling jacket water. These electricity requirements are estimated at 71.4 kJ/kg coal fed to the gasifier (Ref. D-1). Multiplying this factor times the coal feed rate yields

$$71.4 \times 0.528 = \underline{38\ \text{kW}_e\ \text{of electricity}}.$$

The pressure of the low-Btu gas decreases as it passes through the quench/cooling and gas clean-up system. A fan is used to overcome this pressure drop--estimated at 4.98 kPa--and to provide an estimated 7.47 kPa of pressure for distribution of the gas. The energy required by the fan to overcome this pressure drop was calculated by the following equation (Ref. D-2):

$$\begin{array}{l}\text{Electricity} \\ \text{req'd } (\text{kW}_e)\end{array} = 3.03 \times 10^{-5}\ \frac{k}{k-1}\ p_1 q \left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right] (.7457) \left(\frac{1}{E_B E_D}\right) \qquad \text{(Eq. D-1)}$$

where

$k = 1.4$

p_1 = inlet pressure, lb_f/ft^2

p_2 = outlet pressure, lb_f/ft^2

q = gas flow rate, acfm

E_B = efficiency of blower = 60 percent

E_D = efficiency of driver = 90 percent

For the low sulfur coal case,

$$p_1 = 2020 \ \mathrm{lb_f/ft^2}$$
$$p_2 = 2300 \ \mathrm{lb_f/ft^2}$$
$$q = 3979 \ \mathrm{acfm}$$

and the electricity requirements for the fan are calculated at

44 $\mathrm{kW_e}$ of electricity.

Electricity is also required to recirculate quench water to the in-line quench, tray scrubber, and spray scrubber. For the low sulfur coal case, the raw gas is cooled from 250°C (480°F) to 44°C (110°F), and 9.81 kg-moles/hr (21.6 lb-moles/hr) of water are condensed.

In the in-line quench, the gas is saturated and cooled to 64°C (146°F). Cooling duty for the gas is estimated as

$$\frac{0.678 \ \text{kg-moles}}{\text{s}} \times (250 - 64)^\circ\text{C} \times \frac{0.031 \ \text{MW-s}}{\text{kg-mole}^\circ\text{C}} = 0.391 \ \mathrm{MW_T}.$$

This is supplied by evaporation of the 601 kg/hr of water which staurates the gas as it cools.

The enthalpy of the tars was calculated from the following equation (Ref. D-3):

$$H = \frac{1}{1.35} (.302 \ T + .000328 \ T^2 - 19.3) \qquad \text{(Eq. D-2)}$$

where H = enthalpy of tar, Btu/lb
T = temperature, °F

Cooling the 420 lb/hr (0.053 kg/s) of tars from 480°F (110°C) to 146°F (64°C) requires

$$420\ \frac{\text{lb}}{\text{hr}} \times (149 - 24)\ \frac{\text{Btu}}{\text{lb}} \times \frac{\text{MW-hr}}{3.41 \times 10^6\ \text{Btu}} = 0.015\ \text{MW}_T.$$

This is supplied by 1930 kg/hr of quench water which is heated from 55°C (130°F) to 64°C (146°F). Thus, the total quench water to the in-line quench is

$$1930 + 601 = 2530\ \text{kg/hr}.$$

In the plate scrubber, the gas is cooled from 64°C (146°F) to 55°C (130°F). At 55°C, the gas contains 15.13 mole percent water, or 38.1 kg-moles/hr (83.9 lb-moles/hr) water. At 64°C, the gas contains 63.5 kg-moles/hr (140 lb-moles/hr) water. The cooling duty for the water that condenses is therefore

$$\frac{\text{hr}}{3600\ \text{s}} \times 25.4\ \frac{\text{kg-mole}}{\text{hr}} \times \frac{18\ \text{kg}}{\text{kg-mole}} \times \frac{2.38\ \text{MW-s}}{\text{kg}} = 0.302\ \text{MW}_T.$$

Cooling duty for the sensible heat of the gas is estimated at

$$\frac{0.0699\ \text{kg-moles}}{\text{s}} \times (64 - 55)°\text{C} \times \frac{0.0305\ \text{MW-s}}{\text{kg-mole}°\text{C}} = 0.019\ \text{MW}_T.$$

Thus, the total cooling duty for the plate scrubber is

$$0.019 + 0.302 = 0.321\ \text{MW}_T.$$

Quench water is available at 46°C (115°F), and heats to 57°C (135°F). Therefore, the quench water rate to the plate scrubber is

$$0.321\ \text{MW} \times \frac{\text{kg}°\text{C}}{0.00418\ \text{MW-s}} \times \frac{1}{(57-46)°\text{C}} \times \frac{3600\ \text{s}}{\text{hr}} = 25{,}000\ \text{kg/hr}.$$

In the spray scrubber, gas is cooled from 55°C (130°F) to 43°C (110°F). At 43°C, the gas contains 8.67 mole percent H_2O, or 20.3 kg-moles/hr (44.7 lb-moles/hr) water. Thus, 17.8 kg-moles/hr (39.2 lb-moles/hr) water are condensed. Cooling duty for condensation is

$$\frac{\text{hr}}{3600\ \text{s}} \times \frac{17.8\ \text{kg-moles}}{\text{hr}} \times \frac{18\ \text{kg}}{\text{kg-mole}} \times \frac{2.41\ \text{MW-s}}{\text{kg}} = 0.214\ \text{MW}_T.$$

Cooling duty for the sensible heat of the gas is

$$\frac{0.0650\ \text{kg-moles}}{\text{s}} \times (55\text{-}43)^\circ\text{C} \times \frac{0.0305\ \text{MW-s}}{\text{kg-mole}^\circ\text{C}} = 0.024\ \text{MW}_T.$$

Thus, the total cooling duty for the tray scrubber is

$$0.214 + 0.024 = 0.238\ \text{MW}_T.$$

Quench water enters at 35°C (95°F) and exists at 46°C (115°F). Therefore, the quench water rate is

$$0.238\ \text{MW} \times \frac{\text{kg}^\circ\text{C}}{0.00418\ \text{MW-s}} \times \frac{1}{(46\text{-}35)^\circ\text{C}} \times \frac{3600\ \text{s}}{\text{hr}} = 18{,}600\ \text{kg/hr}.$$

From the above calculations, the total quench water circulated to the quench/cooling system is

$$\begin{array}{r} 2{,}530 \\ 25{,}000 \\ \underline{18{,}600} \\ 46{,}100 \end{array} \text{ kg/hr (102,000 lb/hr).}$$

Assuming a liquid head of 0.68 MPa (100 psi), pump efficiency of 55 percent, and driver efficiency of 80 percent, the electricity requirements for the quench liquor pump are calculated as follows:

$$46{,}100\ \frac{\text{kg}}{\text{hr}} \times \frac{\text{hr}}{3600\ \text{s}} \times 0.68\ \text{MPa} \times \frac{\text{kW}\cdot\text{s}}{\text{MPa}\cdot\text{kg}} \times \frac{1}{0.8 \times 0.55} = \underline{20\ \text{kW}_e\ \text{electricity}}.$$

Electricity is required to operate the electrostatic precipitator used for final clean-up of tars, oils, and particulates. The requirements were estimated from Ref. D-4 at:

$\underline{6\ \text{kW}_e\ \text{of electricity}}$.

Energy requirements for the Stretford process were estimated at 1.07 kW_e of electricity per kg of sulfur recovered per hour (Ref. D-5). For the low sulfur coal case, 8.17 kg/hr of sulfur are recovered. Therefore, electricity required by the Stretford process is

$$1.07 \times 8.17 = \underline{9\ \text{kW}_e\ \text{of electricity}}.$$

Triple effect evaporators are used to evaporate excess process condensate. For this example, 177 kg/hr (390 lb/hr) of water are evaporated, and the steam requirements are estimated at

$\underline{0.12\ \text{MW}_T}$.

The results of the calculations just presented are summarized in Table D-1. For the electricity needs, the "real" energy consumed is the thermal energy required to generate the electricity. These values are shown in Table D-1 and were calculated based on an assumed electricity production efficiency of 34.1 percent.

TABLE D-1. SUMMARY OF ENERGY CONSUMPTION - LOW SULFUR COAL CASE

Source of Energy Consumption	Type of Energy Consumed	Energy Consumption, MW_T
Conversion Losses	Coal Feed	1.02
By-Product Tars	Coal Feed	1.96
Inlet Air Blower and Cooling Jacket Pump	Electricity	0.11*
Quench Water Pumps	Electricity	0.06*
Product Gas Fan	Electricity	0.13*
Electrostatic Precipitator	Electricity	0.02*
Stretford Pumps and Fans	Electricity	0.03*
Process Condensate Evaporators	Steam	0.12
Total		3.45

* Electrical energy is based on energy input to the generating source and an assumed conversion efficiency of 34.1 percent.

D.2 HIGH SULFUR COAL/MEA SYSTEMS - EXAMPLE CALCULATION

As mentioned previously, the procedures for calculating energy consumption are the same for all cases examined through the quench/cooling section. Those procedures, as well as the method used for calculating the Stretford process energy requirement were just presented. In the following text, an example calculation is shown for the MEA/Claus/SCOT sulfur removal/recovery system for the high sulfur coal cases.

Energy consumption calculations for the MEA system were based on information from References D-6, D-7, D-8, and D-9. A 20 percent MEA solution was assumed. H_2S sorption was assumed to be the equilibrium value, and CO_2 sorption to be 75 percent of the equilibrium value. Energy requirements for regeneration were assumed to be 310 MJ/m^3 of solution regenerated. Electricity requirements of 630 kW_e per m^3 per second of solution were assumed. A gas pressure of 0.27 MPa psia was used. A residual of 30 ppmv COS remains in the gas, and eighty percent of the COS removed is hydrolyzed to H_2S.

A basis of 8.8 MW_T of gas production was assumed. The compositions of the inlet gas, the outlet gas cleaned to meet the moderate control level, and the acid gas are given in Table D-2. The heating value of the product gas is 6.07 MJ/m^3. Thus, 86.8 m^3/min of gas give 8.8 MW_T. For this gas flow rate, the electricity requirements of the fan required to compress the gas to 0.27 MPa are calculated from Eq. D-1 as

$$\underline{370\ kW_e\ \text{of electricity}}.$$

The MEA solution flow rate is calculated by multiplying the design liquid to gas flow rate of 0.979 m^3/1000 m^3 by the gas flow rate:

$$\frac{0.979\ m^3}{1000\ m^3} \times \frac{86.8\ m^3}{min} \times \frac{min}{60\ s} = 0.00142\ m^3/s.$$

TABLE D-2. GAS COMPOSITIONS FOR MEA SYSTEM, MODERATE CONTROL LEVEL

	Inlet Gas Vol. %	Product Gas Vol. %	Acid Gas Vol. %
CO_2	3.27	1.46	66.9
CO	27.58	28.37	-
CH_4	2.75	2.64	6.5
H_2	13.65	14.04	-
H_2O	3.19	3.28	-
N_2	48.79	50.2	-
H_2S	7424 ppmv	290 ppmv	26.6
COS	297 ppmv	30 ppmv	-
% of Inlet Flow Rate	100%	97.2%	2.77%

Regeneration steam requirements are then equal to

$$0.00142 \frac{m^3}{s} \times \frac{310 \text{ MW-s}}{m^3} = \underline{0.44 \text{ MW}_T}.$$

Finally, electricity requirements are

$$630 \frac{\text{kW-s}}{m^3} \times 0.00142 \frac{m^3}{s} = \underline{0.9 \text{ kW}_e \text{ of electricity}}.$$

The acid gas stream is 2.77 percent of the inlet gas, or 2.4 m^3/min. It contains 1.7 kg-moles/hr of H_2S. Energy requirements for the Claus plant were estimated from data reported for a system treating a gas stream consisting of 40 percent H_2S and 60 percent CO_2 (Ref. D-1, D-10). (The acid gas stream in this example is slightly greater than 26 percent H_2S). The energy requirements are as follows:

Power = 2.8 kWh/kg-mole H_2S.
Steam Credit = 194 kg steam/kg-mole H_2S.

For this example:

Power requirements = $\underline{4 \text{ kW}_e \text{ of electricity}}$.
Steam credit = $\underline{180 \text{ kW}_T}$ (assuming 0.3 MPa steam).

Energy requirements for the SCOT process were estimated by engineering calculations based on data reported in References D-1 and D-11. They are as follows:

Low Pressure Steam = 11.4 kg/kg sulfur (Ref. D-1)
Power = 0.16 kWh/lb sulfur (Ref. D-11)
Fuel Gas = 5.12×10^3 Btu/lb sulfur (theoretical calculation; somewhat greater than reported in Ref. D-1).

For this example, the Claus plant is assumed to achieve 90 percent sulfur recovery. Sulfur recovery in the SCOT is thus 5.22 kg/hr, and energy requirements are:

35 kW_T of steam,
2 kW_e of electricity, and
17 kW_T of fuel gas (low-Btu product gas).

Total energy required for the MEA/Claus/SCOT system is 1100 kW_T of electricity (at an assumed electrical generation efficiency of 34.1 percent), 475 kW_T of steam and 20 kW_T of fuel gas minus 180 kW_T of steam credit from the Claus, or a total of 1390 kW_T (4.74×10^6 Btu/hr).

References

D-1. Dravo Corporation. Handbook of Gasifiers and Gas Treatment Systems, Final Report. Task Assignment No. 4. Pittsburgh, PA. Chemical Plants Division. February 1976.

D-2. Peters, Max S., and Klaus D. Timmerhaus. Plant Design and Economics for Chemical Engineers, second ed. New York. McGraw-Hill. 1968.

D-3. Lowry, H. H. ed. Chemistry of Coal Utilization. Volumes I, II, & Supplementary Volume, New York. Wiley. 1945. 1963. (Supplementary Volume).

D-4. Heinrich, R. F. and J. R. Anderson. Electro-Precipitators. Chemical Engineering Practice, Volume III. H. W. Cremer, ed. New York. Academic. 1975. pp. 484-534.

D-5. Telephone communication between P.J. Murin (Radian Corp.) and Dale Williams and Buz Zey (J.F. Pritchard and Co.) concerning costs of Stretford process. 30 October 1978.

D-6. Maddox, R. N. Gas and Liquid Sweetening. Norman, OK. John M. Campbell Co. 1974.

D-7. Riesenfeld, F. C. and A. C. Kohl. Gas Purification, Second Edition. Houston, Texas. Gulf Publishing Co. 1974.

D-8. Perry, Charles, R. "Basic Design and Cost Data on MEA Treating Units". In Proceedings of the 1967 Gas Conditioning Conference. University of Oklahoma. Norman, OK.

D-9. Kent, Raymond L., and Benjamin Eisenberg. "Better Data for Amine Treating." Hydrocarbon Process. 55(2). 87-90. 1976.

D-10. Goar, Gene. "Impure Feeds Cause Claus Plant Problems." Hydrocarbon Process. 53(7). 129-32. 1974.

D-11. Naber, J. E., J. A. Wesslingh and W. Groenendaal. "New Shell Process Treats Claus Off-Gas." Chemical Engineering Progr. 69(12). 29-34. 1973.